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Economics of Nitrogen Oxides, Sulfur Oxides, and Ash Control Systems for Coal-Fired Utility Power Plants

by

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Economics	Sulfur Dioxide	Stationary Sources	05C	
Coal	Fly Ash	Selective Catalytic Reduction	21D	
Combustion	Catalysis		21B	
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ABSTRACT

A U.S. Environmental Protection Agency (EPA)-sponsored economic evaluation was made of three process combinations to reduce NO_x, SO_2, and ash emissions from coal-fired utility power plants. One case is based on a 3.5% eastern bituminous coal; the other two are based on 0.7% western subbituminous coal. NO_x control is based on an 80% reduction from current new source performance standards (NSPS); SO_2 and fly ash control are based on meeting the current NSPS. Selective catalytic reduction (SCR) is used for NO_x control in all three cases. Limestone scrubbing and a cold-side electrostatic precipitator (ESP) are used in the 3.5% sulfur coal case. Lime spray dryer flue gas desulfurization (FGD) and a baghouse for particulate collection are used in one 0.7% sulfur coal case; limestone scrubbing and a hot-side ESP are used in the other. The economics consist of detailed breakdowns of the capital investments and annual revenue requirements. For systems based on a 500-MW power plant, capital investments range from \$167 to \$187 million (333 to 373 \$/kW) and first-year annual revenue requirements from \$54 to \$60 million (29 to 33 mills/kWh). The 3.5% sulfur coal case is highest because of the higher SO_2 control costs. The case with the spray dryer and baghouse is marginally lower in cost than the case with limestone scrubbing and hot-side ESP. Costs for NO_x control range from one-fourth to one-half of the total costs, largely because of the high cost of the catalyst. The costs of the overall systems and the relationships of the component costs are complexly interrelated because of the interactions of the three processes.

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ABBREVIATIONS AND CONVERSION FACTORS

ABBREVIATIONS

aft3/min	actual cubic feet per minute
Btu	British thermal unit
oF	degrees Fahrenheit
dia	diameter
FGD	flue gas desulfurization
ft	feet
ft2	square feet
ft3	cubic feet
gal	gallon
gpm	gallons per minute
gr	grain
hp	horsepower
hr	hour
in.	inch
k	thousand
kW	kilowatt
kWh	kilowatthour
lb	pound
L/G	liquid-to-gas ratio in gallons per thousand actual cubic feet of gas at outlet conditions
M	million
mi	mile
mo	month
MW	megawatt
ppm	parts per million
psig	pounds per square inch (gauge)
rpm	revolutions per minute
SCA	specific collection area
sec	second
sft3/min	standard cubic feet per minute (60oF)
SS	stainless steel
yr	year

CONVERSION FACTORS

EPA policy is to express all measurements in Agency documents in metric units. Values in this report are given in British units for the convenience of engineers and other scientists accustomed to using the British systems. The following conversion factors may be used to provide metric equivalents.

	To convert British	Multiply by	To obtain Metric	
ac	acre	0.405	hectare	ha
Btu	British thermal unit	0.252	kilocalories	kcal
°F	degrees Fahrenheit minus 32	0.5556	degrees Celsius	°C
ft	feet	30.48	centimeters	cm
ft^2	square feet	0.0929	square meters	m^2
ft^3	cubic feet	0.02832	cubic meters	m^3
ft/min	feet per minute	0.508	centimeters per second	cm/s
ft^3/min	cubic feet per minute	0.000472	cubic meters per second	m^3/s
gal	gallons (U.S.)	3.785	liters	L
gpm	gallons per minute	0.06308	liters per second	L/s
gr	grains	0.0648	grams	g
gr/ft^3	grains per cubic foot	2.288	grams per cubic meter	g/m^3
hp	horsepower	0.746	kilowatts	kW
in.	inches	2.54	centimeters	cm
lb	pounds	0.4536	kilograms	kg
lb/ft^3	pounds per cubic foot	16.02	kilograms per cubic meter	kg/m^3
lb/hr	pounds per hour	0.126	grams per second	g/s
psi	pounds per square inch	6895	pascals (newton per square meter)	Pa (N/m^2)
mi	miles	1609	meters	m
rpm	revolutions per minute	0.1047	radians per second	rad/s
sft^3/min	standard cubic feet per minute (60°F)	1.6077	normal cubic meters per hour (0°C)	m^3/hr (0°C)
ton	tons (short)	0.9072	metric tons	tonne
ton/hr	tons per hour	0.252	kilograms per second	kg/s

ECONOMICS OF NITROGEN OXIDES, SULFUR OXIDES, AND ASH CONTROL SYSTEMS FOR COAL-FIRED UTILITY POWER PLANTS

EXECUTIVE SUMMARY

This is a summary of a U.S. Environmental Protection Agency (EPA)-sponsored evaluation in which the economics of three combinations of emission control processes for coal-fired power plants are examined. Each combination provides for the control of NO_x, SO_2, and fly ash and bottom ash. They represent typical adaptations of current emission control technology as influenced by the types of coal used by utilities, particularly the low-sulfur, predominately western subbituminous coals that have influenced the types of processes used for SO_2 and fly ash control. The first combination, case 1, is based on a high-sulfur eastern bituminous coal. Cases 2 and 3 are based on a low-sulfur western subbituminous coal. The emission control requirements are based on the assumption that a higher degree of NO_x control than now required is necessary; the SO_2 and fly ash control requirements are based on meeting the current 1979 new source performance standards (NSPS).

INTRODUCTION

Most NO_x emission control requirements now in force are being met by modifications to the boiler combustion process. Combustion modifications now being commercially used usually involve reducing the flame temperatures and limiting the amount of oxygen available in the flame zone. These modifications include techniques such as staged combustion (bias firing, burners out of service, and overfire air), flue gas recirculation, low excess air, and dual-register burner designs. Advanced burner and furnace designs now under development have the potential to provide significantly lower NO_x emissions than today's standards. These new combustion systems include fuel-staging and after-burning approaches. However, these new designs are still several years from commercial availability. If stricter regulations were adopted in the near future, these combustion modification methods would not--at least for several years--be adequate and flue gas treatment would be necessary. The most highly developed method of flue gas treatment for NO_x control is selective catalytic reduction (SCR) in which the flue gas is treated with ammonia and passed over a solid catalyst to reduce the NO_x to molecular nitrogen. The need for flue gas treatment to meet NO_x emission limits would probably be met by the use of SCR processes, several variants of which are offered commercially. A generic SCR process, derived from these commercial processes, is therefore used in all three cases.

Limestone scrubbing remains the predominant method of flue gas desulfurization (FGD), increasingly with provisions to produce gypsum by forced or natural oxidation to reduce waste disposal problems. The use of low-sulfur coal has, however, led to the rapid adoption of spray dryer FGD in which the flue gas is contacted with a fine spray of absorbent that evaporates to solid particles in the spray dryer and can be collected as a solid. More than a dozen spray dryer FGD systems have been selected by utilities for low- and medium-sulfur coal applications in the last five years. This trend is represented by the use of a lime-based spray dryer system in case 2, one of the low-sulfur coal cases. For case 1, the high-sulfur coal case, and case 3, the other low-sulfur coal case, conventional limestone FGD systems producing gypsum are used.

The use of low-sulfur coal has also led to the adoption of new methods of fly ash control because the ash is difficult to collect in conventional cold-side (after the boiler air heater) electrostatic precipitators (ESPs) that have served as the industry standard for many years. In many such cases, hot-side (before the air heater) ESPs have been used because the higher ash temperature improves the electrical properties of the ash that affect the efficiency of collection. In both cases, however, strict fly ash emission regulations such as the 1979 NSPS strained the capabilities of then-existing ESP technology, leading to the rapid adoption of fabric filter baghouses for fly ash control. Baghouses, which are proving quite effective, have also been the predominant choice for use with spray dryer FGD in which the fly ash and FGD wastes are collected together. These uses are represented by a conventional cold-side ESP in case 1, a baghouse in case 2, and a hot-side ESP in case 3.

PROCESS DESCRIPTIONS

The base case designs are applied to a new, 500-MW boiler fired with pulverized coal that operates 5,500 hr/yr for 30 years. The boiler meets the 1979 NSPS NO_x emission requirements by combustion modification techniques. The emission control systems are designed for an 80% reduction in these NO_x emissions and for reduction of SO_2 and fly ash emissions to the 1979 NSPS levels. The designs upon which the costs are based include all equipment involved in the collection and disposal of wastes, including a common onsite landfill, and all boiler modifications--air heater modifications and larger induced-draft (ID) fans--made necessary by the presence of the emission control systems. Major conditions are shown in Table S-1.

The SCR systems consist of two trains of insulated reactors with ash hopper bottoms (except in case 3 with an upstream ESP) and provisions for changing catalyst beds. The beds are composed of 0.15- by 0.15- by 1-meter honeycomb blocks in metal modules. Flue gas is ducted from the economizer (or hot-side ESP) outlet and returned to the air heater. Modifications to the air heater to accommodate ammonia salt buildup are included. An ammonia storage and handling system to inject an ammonia-air mixture in the inlet duct is provided. An economizer bypass to maintain the reactor temperature during low-load operation is included. The catalyst life is assumed to be one year, with changes during scheduled boiler outages.

TABLE S-1. MAJOR DESIGN CONDITIONS

	Case 1	Case 2	Case 3
Coal and boiler conditions			
Coal	East. bit.	West. subbit.	West. subbit.
Coal sulfur, % as fired	3.36	0.48	0.48
Coal ash, % as fired	15.1	6.3	6.3
Btu/lb, as fired	11,700	8,200	8,200
Sulfur emitted, % of total	92	85	85
Fly ash, % of total ash	80	80	80
NO_x emitted, lb/MBtu	0.6	0.5	0.5
Boiler size[a], MW	500	500	500
Heat rate, Btu/kWh	9,500	10,500	10,500
Emission control			
NO_x control	SCR	SCR	SCR
NO_x reduction, %	80	80	80
SO_2 control	Limestone FGD	Lime spray dryer	Limestone FGD
Absorber trains[b]	5	4	5
Bypassed flue gas, %	0	12	28
SO_2 removal, overall %	89	65	65
SO_2 removal, absorber %	89	73	90
Fly ash control	Cold-side ESP	Baghouse	Hot-side ESP
Fly ash removal, %[c]	99.7	99.9	99.6

a. Based on coal consumption and heat rate.
b. Including one spare train.
c. In collection device, excluding upstream fallout.

The limestone FGD systems consist of multiple trains of spray tower absorbers connected to a common inlet plenum and discharging to the stack plenum. Each train consists of a presaturator, the absorber with a hold tank and the associated absorbent recirculation system (and an oxidation tank in case 1), and a booster fan. A steam reheater is included in case 1 to provide a stack temperature of 175°F. Flue gas is bypassed in the low-sulfur coal case to eliminate reheating costs (in this case, a less expensive alternative than full scrubbing at the low removal efficiency required). A single slurry preparation area supplies the system. The gypsum waste is dewatered in a thickener and rotary vacuum filters and trucked one mile to the landfill. A spare absorber train and emergency bypasses for one-half of the scrubbed flue gas are provided in all cases. A similar arrangement is used for the spray dryer system in case 2 except that the baghouse booster fans also serve for the spray dryer system. The spray dryers are cylindrical vessels with conical bottoms with single rotary atomizers. The absorbent slurry consists of slaked lime and recycled solids from the baghouse.

The ash control systems consist of the ESPs or baghouses (two parallel identical units), hoppers, conveying systems, a bottom ash dewatering system, storage silos for fly ash, and the equipment for trucking the waste to the landfill. The bottom ash is collected in a conventional hopper and sluiced to the dewatering system. Fly ash is conveyed to silos with a vacuum-pneumatic system. The mixed fly ash and FGD waste in case 2 are conveyed with a pressure-vacuum pneumatic system.

ECONOMIC PROCEDURES

The economics consist of the capital investment in 1982 dollars and the first-year and levelized annual revenue requirements in 1984 dollars. The annual revenue requirements consist of operating and maintenance costs plus capital charges. The capital charges are levelized in both the first-year and levelized annual revenue requirements; whereas, the operating and maintenance costs are also levelized in the latter. The levelizing factor in all cases is 1.886, which represents a 6% annual inflation and a 10% discount rate over the 30-year life of the project.

The costs include all costs associated with the construction and operation of the systems, including modifications to the boiler air heater and the incremental increase in the boiler ID fans to account for the pressure drop in the emission control equipment that is not compensated for by separate booster fans. The construction and operating costs of the landfill are also included.

The costs are divided into three sections representing costs for NO_x, SO_2, and ash control and are further divided into categories representing particular unit operations within the processes. In cases in which equipment or operations serve more than one process (incremental increases in boiler ID fans and the common landfill, for example), the costs are prorated using the appropriate factors (pressure drops or waste volume, for example). Baghouse costs are not prorated, however, because of the effect of flue gas volume on baghouse costs.

RESULTS

The capital investments and annual revenue requirements are summarized in Tables S-2 through S-5. With the choice of processes determined, at least in part, by the type of coal, and the costs of the individual processes influenced by other processes in the system, economic comparisons on a process-by-process basis must be interpreted with care, as seen in the detailed breakdown of the base case costs.

TABLE S-2. SUMMARY OF CAPITAL INVESTMENTS IN M$[a,b]

	Capital investment, mid-1982 $ M$			
	NO_x	SO_2	Particulate	Total
Base case, 500 MW, 80% NO_x removal				
Case 1	41.9	101.8	42.9	186.6
Case 2	50.1	54.0	62.6	166.7
Case 3	48.1	69.4	53.5	171.0
Case variation, 200 MW, 80% NO_x removal				
Case 1	20.6	58.2	22.6	101.5
Case 2	24.2	31.7	31.4	87.3
Case 3	24.3	41.4	27.8	93.5
Case variation, 1,000 MW, 80% NO_x removal				
Case 1	77.7	175.7	73.3	326.6
Case 2	94.8	97.4	110.7	302.9
Case 3	91.2	121.1	94.6	306.9
Case variation, 500 MW, 90% NO_x removal				
Case 1	48.2	101.9	42.9	193.0
Case 2	55.5	54.0	62.7	172.2
Case 3	53.9	69.4	53.5	176.8

a. Table S-1 lists the major design conditions for each case.

b. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

TABLE S-3. SUMMARY OF CAPITAL INVESTMENTS IN \$/KW[a,b]

	Capital investment, mid-1982 \$ \$/kW			
	NO_x	SO_2	Particulate	Total
Base case, 500 MW, 80% NO_x removal				
Case 1	83.7	203.7	85.8	373.2
Case 2	100.2	108.0	125.3	333.4
Case 3	96.1	138.7	107.1	342.0
Case variation, 200 MW, 80% NO_x removal				
Case 1	103.1	291.0	113.2	507.3
Case 2	121.0	158.3	157.2	436.6
Case 3	121.6	206.9	139.0	467.5
Case variation, 1,000 MW, 80% NO_x removal				
Case 1	77.7	175.7	73.3	326.6
Case 2	94.8	97.4	110.7	302.9
Case 3	91.2	121.1	94.6	306.9
Case variation, 500 MW, 90% NO_x removal				
Case 1	96.4	203.8	85.8	386.0
Case 2	111.0	108.0	125.4	344.3
Case 3	107.8	138.8	107.1	353.6

a. Table S-1 lists the major design conditions for each case.
b. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

Base Case Capital Investments

Breakdowns of the base case capital investments are shown in Table S-6. The case 1 (3.5% sulfur coal, SCR, limestone FGD, and cold-side ESP) capital investment is \$187 million (373 \$/kW), of which NO_x control accounts for 22% of the total; SO_2 control, 55%; and particulate control, 22%. The case 2 (0.7% sulfur coal, SCR, spray dryer FGD, and baghouse) capital investment is \$167 million (333 \$/kW) and the breakdown is 30%, 32%, and 38%. The case 3 (0.7% sulfur coal, hot-side ESP, SCR, and limestone FGD) capital investment is \$171 million (342 \$/kW) and the breakdown is 28%, 40%, and 32%. The low percentage for SO_2 control in case 2 with the spray dryer results from the

TABLE S-4. SUMMARY OF ANNUAL REVENUE REQUIREMENTS IN M$[a,b]

	Annual revenue requirements, 1984 $							
	First year				Levelized			
	M$				M$			
	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total
Base case, 500 MW, 80% NO_x removal								
Case 1	21.9	28.8	9.8	60.4	35.8	41.0	12.8	89.7
Case 2	26.5	12.7	14.4	53.6	43.5	16.9	19.0	79.4
Case 3	24.7	18.0	12.1	54.8	40.4	24.9	15.8	81.0
Case variation, 200 MW, 80% NO_x removal								
Case 1	9.7	16.3	5.2	31.2	15.6	23.1	6.9	45.6
Case 2	11.6	7.6	7.7	26.8	18.7	10.1	10.4	39.2
Case 3	11.1	11.0	6.6	28.7	17.8	15.3	8.8	41.9
Case variation, 1,000 MW, 80% NO_x removal								
Case 1	41.5	48.8	16.1	106.4	68.1	69.2	20.9	158.2
Case 2	51.2	22.2	24.5	97.9	84.2	29.2	31.8	145.1
Case 3	47.9	30.3	20.5	98.7	78.4	41.4	26.4	146.3
Case variation, 500 MW, 90% NO_x removal								
Case 1	26.1	28.8	9.8	64.6	42.9	41.0	12.8	96.7
Case 2	30.1	12.7	14.4	57.2	49.5	16.9	19.0	85.4
Case 3	28.6	18.0	12.1	58.6	46.8	24.9	15.8	87.5

a. Table S-1 lists the major design conditions for each case.

b. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

TABLE S-5. SUMMARY OF ANNUAL REVENUE REQUIREMENTS IN MILLS/KWH[a,b]

	Annual revenue requirements, 1984 $							
	First year				Levelized			
	Mills/kWh							
	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total
Base case, 500 MW, 80% NO_x removal								
Case 1	8.0	10.5	3.5	22.0	13.0	14.9	4.7	32.6
Case 2	9.6	4.6	5.2	19.5	15.8	6.2	6.9	28.9
Case 3	9.0	6.5	4.4	19.9	14.7	9.0	5.7	29.5
Case variation, 200 MW, 80% NO_x removal								
Case 1	8.8	14.8	4.7	28.4	14.2	21.0	6.3	41.5
Case 2	10.6	6.9	7.0	24.4	17.0	9.2	9.4	35.7
Case 3	10.1	10.0	6.0	26.1	16.2	13.9	8.0	38.1
Case variation, 1,000 MW, 80% NO_x removal								
Case 1	7.5	8.9	2.9	19.3	12.4	12.6	3.8	28.8
Case 2	9.3	4.0	4.5	17.8	15.3	5.3	5.8	26.4
Case 3	8.7	5.5	3.7	18.0	14.3	7.5	4.8	26.6
Case variation, 500 MW, 90% NO_x removal								
Case 1	9.5	10.5	3.5	23.5	15.6	14.9	4.7	35.2
Case 2	10.9	4.6	5.2	20.8	18.0	6.2	6.9	31.1
Case 3	10.4	6.5	4.4	21.3	17.0	9.0	5.7	31.8

a. Table S-1 lists the major design conditions for each case.

b. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

TABLE S-6. BASE CASE CAPITAL INVESTMENT COMPARISON[a]

Process capital	Case 1, k$				Case 2, k$				Case 3, k$			
	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total
NH_3 storage and injection	1,314			1,314	1,328			1,328	1,297			1,297
Reactor	7,829			7,829	9,278			9,278	8,453			8,453
Flue gas handling	3,843	11,343	1,311	16,497	4,543	7,374	4,961	16,878	5,386	11,175	4,290	20,851
Air heater	819			819	1,220			1,220	861			861
Materials handling		2,528		2,528		1,132		1,132		1,266		1,266
Feed preparation		4,717		4,717		1,258		1,258		2,363		2,363
SO_2 absorption		20,411		20,411		12,992		12,992		18,070		18,070
Oxidation		2,677		2,677								
Reheat		3,653		3,653								
Solids separation		3,681		3,681								
Lime particulate recycle						2,140		2,140		2,265		2,265
Particulate removal and storage			10,509	10,509			15,446	15,446			14,354	14,354
Particulate transfer			5,636	5,636			6,779	6,779			4,378	4,378
Total process capital, k$	13,805	49,010	17,456	80,271	16,369	24,896	27,186	68,451	15,997	35,139	23,022	74,158
Other Capital Investment												
Waste disposal direct investment	19	4,011	3,344	7,374	34	527	2,749	3,310	30	847	2,628	3,505
Land	10	458	377	845	15	75	326	416	15	113	313	441
Catalyst	12,028			12,028	14,678			14,678	13,455			13,455
Royalty	463			463	563			563	563			563
Other[b]	15,530	48,360	21,710	85,600	18,431	28,478	32,388	79,297	18,001	33,272	27,583	78,856
Total	41,855	101,839	42,887	186,581	50,090	53,976	62,649	166,715	48,061	69,371	53,546	170,978
Total, $/kW[c]	83.7	203.7	85.8	373.2	100.2	108.0	125.3	333.4	96.1	138.7	107.1	342.0

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a. Table S-1 lists the major design conditions for each case.

b. Consists of costs for "services, utilities, and miscellaneous"; all six items of "indirect investment"; "allowance for startup and modifications"; "interest during construction"; and "working capital" as listed in the appendix tables.

c. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

particulate collection costs for FGD waste being combined with the fly ash collection costs and assigned to particulate control costs.

Nitrogen Oxides Control--

For NO_x control, the most important capital cost is the initial catalyst charge, which is almost one-third of the total capital investment. Most of the remaining capital costs are for the reactor and the associated internal and external catalyst supports and handling system, and for the incremental fan cost and flue gas ductwork associated with flue gas handling. The remaining capital costs--ammonia storage and injection system, air heater modification, waste disposal (of spent catalyst), land, and royalties--are relatively minor. Incremental fan costs are minor; 90% of the flue gas-handling costs is for ductwork.

Most of the capital costs are directly related to the flue gas volume, particularly for the major cost areas. As a result, the total capital investment for NO_x control in case 1 is lowest because of lower flue gas volume with the high-Btu coal. Case 3 is slightly lower than case 2 because of the absence of fly ash.

Air heater modification costs are associated with the increase in size, the more tightly packed elements, and the use of thicker and more corrosion-resistant elements.

The ammonia storage and injection costs are almost the same for all three cases. The only cost differences result from differences in the injection grid, which vary with the flue gas duct size and design.

Sulfur Dioxide Control--

The capital investments for SO_2 control are highest for case 1 and lowest for case 2, but the capital investment for case 2 does not contain the costs for FGD waste collection. In all three cases, most of the costs are associated with the SO_2 absorption area (the absorbers and the absorbent liquid system or the spray dryers) and the flue gas-handling area (fans and ductwork). These two areas account for 65% of the process equipment costs in case 1 and about 80% of the process equipment costs in cases 2 and 3.

The higher capital investment for case 1 as compared with case 3 is almost entirely related to the larger quantities of SO_2 removed. The materials-handling (limestone), feed preparation, and solids separation area costs are roughly two times higher and waste disposal costs are almost five times higher for case 1 than for case 3. In addition, the SO_2 removal requirements in case 1 require both full scrubbing--necessitating steam reheat of the flue gas--and forced oxidation, neither of which is necessary in case 3.

SO_x control in case 2 is the least expensive, primarily because of lower costs in the SO_2 absorption area (because there is no liquid recirculation system) and in the flue gas-handling areas (because of the lower pressure drop in the spray dryers and the economy of scale with fan costs prorated between SO_2 and particulate control). An accurate comparison of SO_2 control capital investment in cases 2 and 3, however, must include the costs of particulate collection, which is discussed in the following section.

Particulate Control--

The capital investments for particulate control are $43 million for case 1, $63 million for case 2, and $54 million for case 3. In all three cases, the particulate removal and storage area accounts for about 60% of the total particulate control process equipment costs, with the ESPs or baghouses and their hoppers accounting for most of the area cost. The cold-side ESPs of case 1 have an installed cost of $5.9 million and the hot-side ESPs of case 3 have an installed cost of $9.8 million. Most of this difference is a result of the larger flue gas volume in case 3--both in an absolute sense and because the ESPs in case 3 operate at a higher temperature. (An SCA of 500 $ft^2/kaft^3/min$ was used for case 1 after determining SCA values ranging from about 450 to over 650 $ft^2/kaft^3/min$ from several references. Some reviewers state that an SCA range of 200 to 250 $ft^2/kaft^3/min$ is adequate to meet the ash removal required by the ESP in case 1. If an ESP designed with an SCA of 250 was used, the investment and revenue requirements for particulate control would be reduced about 15%.) The baghouses have an installed cost of $7.4 million. Much of the cost difference between cases 2 and 3 is a result of the larger size of the baghouses and the corresponding larger and more complex hoppers required.

Particulate transfer process equipment costs are $5.6 million for case 1, $6.8 million for case 2, and $4.4 million for case 3. Case 2 has a more complicated pressure-vacuum conveying system, which accounts for most of the cost difference between cases 2 and 3.

Flue gas-handling costs are $1.3 million for case 1, $5.0 million for case 2, and $4.4 million for case 3. The lower costs for case 1 result from the smaller absolute volume and lower temperature of the flue gas. In addition, the costs for cases 1 and 3 are almost totally composed of the cost of ductwork since the incremental fan costs are negligible. In the case of the baghouses, however, fan costs are significant, about equal to ductwork costs, because of the large pressure drop through the baghouses.

Base Case Comparisons - Annual Revenue Requirements

The base case annual revenue requirements are shown in Table S-7. The first-year annual revenue requirements for case 1 (3.5% sulfur coal, SCR, limestone FGD, and cold-side ESP) are $60 million (22 mills/kWh) with 36% associated with NO_x control, 48% with SO_2 control, and 16% with particulate control. For case 2 (0.7% sulfur coal, SCR, spray dryer FGD, and baghouse), the first-year annual revenue requirements are $54 million (19.5 mills/kWh) with 49% associated with NO_x control, 24% with SO_2 control, and 27% with particulate control. For case 3 (0.7% sulfur coal, hot-side ESP, SCR, and limestone FGD), the first-year annual revenue requirements are $55 million (19.9 mills/kWh) with 45% associated with NO_x control, 33% with SO_2 control, and 22% with particulate control.

The levelized annual revenue requirements are $90 million (33 mills/kWh), $79 million (29 mills/kWh), and $81 million (30 mills/kWh) for cases 1, 2, and

TABLE S-7. ANNUAL REVENUE REQUIREMENT ELEMENT ANALYSIS FOR BASE CASES

500-MW UNIT WITH 80% NO_x REMOVAL[a]

Direct costs	Case 1				Case 2				Case 3			
	NO	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total
Ammonia	364			364	336			336	336			336
Catalyst	13,899			13,889	16,962			16,962	15,549			15,549
Lime/limestone		1,216		1,216		708		708		186		186
Operating labor and supervision												
Process	66	658	230	954	66	263	296	625	66	594	230	890
Landfill	3	523	436	962	5	83	435	523	4	127	393	524
Steam	51	1,369		1,420	65			65	63			63
Electricity	278	2,146	581	3,005	492	780	966	2,238	391	1,477	993	2,861
Fuel	1	162	135	298	1	18	95	114	1	28	87	116
Maintenance	586	4,276	1,025	5,887	695	1,599	1,811	4,105	679	3,005	1,299	4,983
Analysis	46	104	6	156	46	88	6	140	46	69	6	121
Other	13	27	19	59	17	16	36	69	41	19	36	96
Total direct costs, k$	15,307	10,481	2,432	28,220	18,685	3,555	3,645	25,885	17,176	5,505	3,044	25,725
Indirect costs												
Overheads	421	3,337	1,018	4,776	487	1,220	1,529	3,236	477	2,277	1,157	3,911
Capital charges	6,153	14,970	6,304	27,427	7,363	7,934	9,209	24,506	7,065	10,198	7,871	25,134
Total first-year annual revenue requirements												
k$	21,881	28,788	9,754	60,423	26,535	12,709	14,383	53,627	24,718	17,980	12,072	54,770
Mills/kWh[b]	8.0	10.5	3.5	22.0	9.6	4.6	5.2	19.5	9.0	6.5	4.4	19.9
Levelized annual revenue requirements												
k$	35,816	41,031	12,811	89,658	43,521	16,940	18,967	79,428	40,359	24,875	15,794	81,028
Mills/kWh[b]	13.0	14.9	4.7	32.6	15.8	6.2	6.9	28.9	14.7	9.0	5.7	29.5

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a. Table S-1 lists the major design conditions for each case.

b. All values have been rounded; therefore, totals do not necesarily correspond to the sum of the individual values indicated.

3, respectively. For cases 1, 2, and 3, respectively, 40%, 55%, and 50% of the total levelized annual revenue requirements are associated with NO_x control; 46%, 21%, and 31% with SO_2 control; and 14%, 24%, and 19% with particulate control.

The cost per ton of pollutant removed is presented for the base cases in Table S-8 based on each of first-year and levelized annual revenue requirements. A comparison on this basis indicates that NO_x control is significantly less cost effective than SO_2 and particulate control. For example, with first-year annual revenue requirements, the costs in Table S-8 range from about 3,500 \$/ton to 4,600 \$/ton for NO_x control, from about 500 \$/ton to over 1,900 \$/ton for SO_2 control, and from 60 \$/ton to 130 \$/ton for particulate control.

TABLE S-8. COST PER TON OF POLLUTANT REMOVED FOR BASE CASES

500-MW UNIT WITH 80% NO_x REMOVAL

	\$/ton, 1984 \$					
	First year			Levelized		
	NO_x	SO_2	Particulate	NO_x	SO_2	Particulate
Case 1	3,490	470	60	5,710	670	80
Case 2	4,600	1,370	130	7,540	1,820	170
Case 3	4,280	1,930	110	6,990	2,680	140

Nitrogen Oxides Control--

The first-year annual revenue requirements for the NO_x control processes in cases 1, 2, and 3, respectively, are \$22 million (8 mills/kWh), \$27 million (10 mills/kWh), and \$25 million (9 mills/kWh). In all cases, the catalyst replacement costs are the overwhelmingly dominant cost elements: over 90% of the direct costs and two-thirds of the total annual revenue requirements are for the yearly replacement of catalyst. Except for this cost, the annual revenue requirements are modest, appreciably less than the costs for similar cost categories for SO_2 and particulate control.

Sulfur Dioxide Control--

The first-year annual revenue requirements for the SO_2 control processes are \$29 million (11 mills/kWh), \$13 million (5 mills/kWh), and \$18 million (7 mills/kWh) for cases 1, 2, and 3, respectively. Again, case 2 with the spray dryer does not include costs associated with operation of the baghouse. Excluding capital charges (which are proportional to capital investment) and overheads (which are proportional to the direct costs), the direct costs of the annual revenue requirements reflect appreciably wider differences in operating costs. The direct costs are \$10.5 million, \$3.6 million, and \$5.5 million for cases 1, 2, and 3, respectively. Maintenance costs are the highest element of direct costs in all three cases, followed again in all three cases by electricity costs. Steam for reheating the flue gas is the third largest direct cost (13% of the total) in case 1, a cost not incurred by cases 2 and 3, which have bypass reheat. These costs and the remaining direct costs are all higher for case 1 than the corresponding costs for cases 2 and 3, a result of the large quantity of SO_2 removed for case 1. With the exception of lime costs, which are 20% of the total direct costs, case 2 has lower direct costs in every category as compared with case 3.

Particulate Control--

The first-year annual revenue requirements for particulate control are \$10 million (4 mills/kWh), \$14 million (5 mills/kWh), and \$12 million (4 mills/kWh) for cases 1, 2, and 3, respectively. The annual revenue requirements for case 2, however, also include the collection of the spray dryer FGD solids. Among the direct costs, maintenance costs are the highest direct cost in all three cases, followed by electricity costs and labor costs. Maintenance costs are highest for case 2, which are about 75% higher than case 1 and 40% higher than case 3. Electricity costs are lowest for case 1 and highest for case 3, while case 2 has only slightly lower electricity costs than case 3. Labor costs do not differ appreciably, although process labor in case 2 is about 25% higher than in cases 1 and 3.

Energy Requirements

The energy consumptions of the base cases, expressed in Btu equivalents, are shown in Table S-9. The total energy requirements range from 4.89% of the boiler capacity for case 1 to 2.31% of the boiler capacity for case 2. The NO_x control energy requirements are the lowest in all three cases and most are for the incremental electricity consumption of the boiler ID fan that compensates for the relatively small pressure loss in the reactors. For SO_x control, cases 1 and 3 have large electricity requirements because of the FGD booster fans and the pumping requirements for the absorbent liquid recirculation systems. These are similar in both cases. The electricity requirements for the spray dryer in case 2 are lower because there is no liquid recirculation system. Particulate control energy requirements in cases 1 and 3 are mostly for ESP electricity, which is substantially lower for the cold-side ESP. In case 2, most of the electricity is for the booster ID fans that compensate for the relatively high pressure drop in the baghouse.

Power Unit Size Case Variation

The capital investments and annual revenue requirements of systems for 200-MW, 500-MW, and 1,000-MW systems are shown in Tables S-2 through S-5. Compared with the 200-MW systems, the 500-MW systems are 83% to 91% higher and the 1,000-MW systems are 222% to 247% higher in capital investment. In terms of \$/kW, the 1,000-MW systems are about one-third less expensive, however, because of the economy of scale. The general relationships of the three cases remain the same at all three power unit sizes. The rate of capital investment increase is greatest for the NO_x control processes (an increase of 275% to 292% between the 200-MW and 1,000-MW sizes, as compared with 193% to 207% for the SO_2 control processes and 224% to 253% for the particulate control processes) and it is also higher for the spray dryer FGD process and the baghouse than for the limestone FGD process and ESPs. As a result, the rate of capital investment increase with size is greatest for case 2.

TABLE S-9. COMPARISON OF BASE CASE ENERGY REQUIREMENTS

Case	Steam, MBtu/hr	Electricity, MBtu/hr	Diesel fuel, MBtu/hr	Percent of power unit, input energy
Case 1[a]				
NO_x	3.15	12.97	0.01	0.34
SO_x	83.79	100.20	2.65	3.93
Particulate	0.00	27.14	2.20	0.62
Total	86.94	140.31	4.86	4.89
Case 2[b]				
NO_x	4.00	25.40	0.02	0.56
SO_x	0.00	40.26	0.30	0.77
Particulate	0.00	49.85	1.55	0.98
Total	4.00	115.51	1.87	2.31
Case 3[b]				
NO_x	3.88	20.18	0.02	0.46
SO_x	0.00	76.20	0.46	1.46
Particulate	0.00	51.22	1.41	1.00
Total	3.88	147.60	1.89	2.92

Note: Does not include energy requirement represented by raw materials.

a. Based on a 500-MW boiler, a gross heat rate of 9,500 Btu/kWh for generation of electricity, and a boiler efficiency of 90% for generation of steam.

b. Based on a 500-MW boiler, a gross heat rate of 10,500 Btu/kWh for generation of electricity, and a boiler efficiency of 90% for generation of steam.

Compared with the 200-MW systems, the annual revenue requirements of 500-MW systems are 91% to 100% higher, the 1,000-MW systems are 241% to 265% higher, and there is approximately a one-third reduction in costs in terms of $/kWh. As with capital investment, the annual revenue requirements retain the same general relationships at the three power unit sizes and the rates of increase for the NO_x control processes are higher (328% to 341% between the 200-MW and the 1,000-MW sizes, compared with 175% to 199% for the SO_2 control processes and 210% to 218% for the particulate control processes) and the rates for the spray dryer FGD and baghouse are higher than those of the limestone FGD systems and ESPs.

Two-Year Catalyst Life Case Variation

To illustrate the effect of catalyst life on annual revenue requirements, the annual revenue requirements for the three 500-MW base cases were also determined for a 2-year catalyst life. The only change in NO_x control annual revenue requirements is a reduction in the catalyst cost by 50%--\$7.0 million, \$8.5 million, and \$7.8 million for cases 1, 2, and 3, respectively. The longer catalyst life reduces the annual revenue requirements of NO_x control by one-third. The annual revenue requirements of the overall systems are reduced by 12% to 16%.

Ninety Percent Nitrogen Oxide Reduction Case Variation

To evaluate the economic effects of a 90% reduction in NO_x, as compared with the 80% used in the other evaluations, the economics of the three 500-MW cases were determined with 90% NO_x reduction. The primary differences from the base case conditions are an NH_3:NO_x ratio of 0.91:1.0 instead of 0.81:1.0, a 12% increase, and an increase in catalyst (based on vendor recommendations) of 22.5% for case 1, 15.0% for case 2, and 18.0% for case 3. The capital investments of the NO_x control processes are increased 11% to 15% and the total for the three systems by 3% to 4%, all of which is a result of the increase in NO_x reduction. The first-year annual revenue requirements for the NO_x process are increased 19%, 14%, and 16% for cases 1, 2, and 3, respectively. The effect on the annual revenue requirements of the overall system of increasing the NO_x from 80% to 90% is an increase of 7% in all three cases.

Ammonia Price Case Variation

Changes in the price of ammonia would have little effect on the overall cost of the NO_x control process. The annual revenue requirements for the NO_x control processes (in the 500-MW base case) increase only 1.5% to 1.9% as the ammonia price is doubled from the base case value of 155 \$/ton to 310 \$/ton.

CONCLUSIONS

The total costs for case 1, based on 3.5% sulfur coal, and cases 2 and 3, based on 0.7% sulfur coal, differ less than 15% in capital investment and annual revenue requirements in spite of the differing control processes. This is a result in part of offsetting differences--the much higher SO_2 control costs for case 1 are offset by lower fly ash control costs and a smaller flue gas volume. The costs for the two low-sulfur coal cases, one with a spray dryer FGD system and baghouse and the other with limestone FGD and a hot-side ESP, differ only marginally in cost. In the two low-sulfur coal cases, the low spray dryer FGD costs and the advantage of combined particulate collection are offset by the higher NO_x control costs and higher baghouse costs. When only the SO_2 and fly ash control costs are compared, the spray dryer-baghouse case is 5% lower in capital investment and 12% lower in annual revenue requirements than the hot-side ESP and limestone FGD case.

The combined emission control processes increase the power plant capital investment by about 35% on the average, of which the NO_x portion is about one-third. Based on levelized annual revenue requirements, the average increase in the cost of power is about 45%, of which the NO_x portion is about one-half.

The energy requirements of 2% to 5% of the boiler input energy are mostly for SO_2 and particulate control. For the cases with limestone FGD, SO_2 control has the highest energy requirements.

The use of flue gas treatment for NO_x control, such as the SCR process in this study, would add significantly to emission control costs. An SCR process for a 500-MW power plant would have a capital investment of 80 to 100 $/kW and annual revenue requirements of 8 to 9 mills/kWh. The high cost is largely associated with the catalyst replacement cost, which accounts for 90% of the direct costs in annual revenue requirements. A 2-year catalyst life reduces the annual revenue requirements by over one-third, however, so the costs for NO_x control in this study, which are based on a 1-year life, could be substantially reduced if extended catalyst lives are attained.

Other than catalyst life, the main factor affecting NO_x control costs is the flue gas volume which determines the fan and ductwork costs and the catalyst volume. Increasing the NO_x reduction efficiency from 80% to 90% increases the costs by 10% to 20%, again because of the larger catalyst volume needed. Ammonia costs have almost no effect on costs; doubling the price of ammonia increases the annual revenue requirements by about 2%.

Although the costs of NO_x control are in the same general range as those for SO_2 and fly ash control, if the processes are compared on the basis of the pounds of pollutants reduced, the costs for NO_x control are 2 to 10 times greater than for SO_2 control and 40 to 60 times greater than for ash control.

In SO_2 control, the major costs are associated with the absorption area and flue gas handling (ductwork and fans). These costs do not differ greatly among the three cases because of offsetting differences--a larger cost for liquid circulation in the high-sulfur coal case but a larger flue gas volume in the low-sulfur coal cases, which requires larger equipment and has larger fan costs. The higher costs for the high-sulfur coal case are in large part the result of the much larger quantity of sulfur removed: the materials-handling, waste-handling, and disposal costs are two to five times higher for the high-sulfur coal case than for the low-sulfur coal case with limestone FGD.

INTRODUCTION

This report describes a U.S. Environmental Protection Agency (EPA)-sponsored economic evaluation of three combinations of flue gas emission control processes, each of which provides for the control of nitrogen oxides (NO_x), for flue gas desulfurization (FGD), and for fly ash removal. The process designs are based on electric utility boilers fired with pulverized coal, using a 500-MW boiler as the base case. The NO_x emission level used as the design basis is substantially lower than the present allowable limits in most areas, representing possible regulatory changes that could result from concerns about the effects of NO_x on large-scale atmospheric phenomena such as haze and acid rain (1). The SO_2 and fly ash emission levels are based on the limits imposed by the 1979 new source performance standards (NSPS) (2).

The processes in each of the three cases represent some of the current trends in emission control technology that are the product of changing patterns of coal use by utilities, responses to existing or anticipated emission control requirements, and of developments that have produced more economical and environmentally acceptable emission control processes. All three cases include selective catalytic reduction (SCR) flue gas treatment for NO_x control. The first case (case 1) is based on the use of a high-sulfur eastern bituminous coal; it includes a forced-oxidation limestone FGD process and a conventional cold-side electrostatic precipitator (ESP) in addition to the SCR process. The other two cases are based on the use of a low-sulfur western subbituminous coal. One (case 2) includes a spray dryer FGD system with a fabric filter baghouse that serves to collect both FGD waste and fly ash. The other (case 3) includes a limestone FGD system (with natural oxidation to gypsum) and a hot-side ESP.

Except for a few areas in the United States, the reductions in NO_x emissions required thus far for utility boilers have been met by modifications to the combustion process that reduce the formation of NO_x in the furnace. These present commercial methods do not appear capable of providing substantially higher levels of NO_x reduction. New burner and furnace designs now under development have the potential to provide significantly lower NO_x emissions than today's standards. However, these promising new combustion modification designs are still several years from commercial availability (3). Flue gas treatment would probably be necessary to meet near-term regulations requiring reductions substantially below the existing 1979 NSPS limits. Numerous flue gas treatment processes to control NO_x emissions have been brought to various stages of development and use, primarily in Japan where large reductions in NO_x emissions are required (4). Among these processes, SCR processes have proven popular because of their simplicity and effectiveness; they have also been evaluated on a prototype scale in the United States (5,6,7,8). SCR processes appear the most likely candidates for widespread use if further large reductions in NO_x emissions become necessary in the near future. Because of this, an SCR process is used in each of the three cases evaluated in this study.

The FGD and particulate control processes represent trends in emission control that have been accelerated during the past few years by the increasing use of low-rank western coals, which are growing in importance as a fuel for electricity generation in the United States (9). Low-rank western coals--typified in quantity of reserves and extent of use by the subbituminous coals of the northern Great Plains and Rocky Mountains (10)--are characterized, as compared with eastern coals, by a low heating value, a low sulfur and ash content, a high moisture content, and a different ash chemical composition. Many of the deposits can be efficiently and economically mined in large volumes (11). Along with efficient coal transportation systems, this has made them economically competitive over a wide area of the central and east-central United States. They have been a source of low-sulfur coal to meet SO_2 emission requirements for some time and are also coming into increasing use in situations that require FGD. This use has been accelerated by the increasing use of coal to generate electricity in the trans-Mississippi west, as well as the attractiveness of these coals as a reliable and economical source of coal east of the Mississippi River (12).

The low-sulfur content of western coals has encouraged the development of alternate methods of FGD that do not involve wet scrubbing. One result of these efforts has been the development and rapid adoption of spray dryer FGD, which is used in one of the three cases in this study. Wet limestone FGD, used in the other two cases, remains widely used in high-sulfur coal applications and is also common in low-sulfur coal applications. The forced-oxidation version used for the high-sulfur coal case in this study represents an increasing use of this process innovation to reduce waste dewatering and disposal problems.

The high-sulfur coal case has a conventional cold-side ESP, a type traditionally used in this type of application and one which has long provided reliable and economic fly ash removals in the high 90% range. The use of low-sulfur coals, however, created collection problems (the removal efficiency of cold-side ESPs depends in large part on the presence of conditioners such as SO_3 in the flue gas) that led to the widespread use of hot-side ESPs, and more recently, to fabric filter baghouses. Both types of ESPs remain in widespread use in new construction but the difficulties of attaining the very high removal efficiencies necessary for some emission regulations--exacerbated to some degree by the use of western coals, which not only have a low-sulfur content but an ash composition inimical to collection by electrostatic methods--have led to an increasing use of fabric filter baghouses. The frequent use of fabric filters with spray dryer FGD has also increased the acceptance of baghouses.

The processes used in this study represent several aspects of emission control technology: The SCR process is a promising method of meeting low limits of NO_x emissions unattainable with the methods now in use; it is, however, unlikely to replace these methods to meet existing emission limits on its own merits. Hot-side ESPs and fabric filter baghouses are responses to the problems encountered with some coals in attaining low fly ash emission limits with conventional cold-side ESPs. Spray dryer FGD, on the other hand, is an attractive alternative to conventional wet scrubbing and is replacing wet scrubbing in some applications. Finally, the use of forced oxidation in the limestone FGD process is a response not related to emission control, but to problems of waste handling and disposal.

BACKGROUND

All of the processes used in this study have been extensively discussed in the technical literature. The following discussion is primarily a perspective of this literature with emphasis on details that have a direct bearing on the processes in relation to the coals used and their interactions in various combinations that have economic effects. Some of the processes--notably the SCR process and the hot-side ESP--also intrude on boiler functions in ways that directly affect the boiler operation.

The SCR process (13), the limestone FGD process (14), and spray dryer FGD (15) have been evaluated in previous Tennessee Valley Authority (TVA) economic studies for EPA. Ash-handling and disposal economics--excluding the actual collection of the ash--have also been evaluated in a similar study (16). These evaluations focused on the economic comparison of individual processes over a range of conditions.

The type of boiler also has an effect on the economics of some of the processes. In terms of number and generating capacity, a dry-bottom boiler fired with pulverized coal is most typical of utility boilers. About three-fourths of the coal used by electric utilities is burned in this type of boiler (17) and the type also predominates in new construction (18). In this type of boiler, the coal is reduced to a fine powder and blown into the furnace as a suspension in part of the combustion air. The term dry bottom is applied to designs in which the ash solidifies as small particles while suspended in the combustion gases, a part of which falls from the bottom of the furnace as "dry" bottom ash. Similar wet-bottom furnaces are designed so that the ash collects as molten slag on the furnace walls and drains from the furnace in molten form. Wet-bottom designs are usually used only when problem fuels make a dry-bottom design impractical. The other types of coal-fired boilers used by utilities are stoker fired (which, although numerous, are small and consume only about 1% of the coal used by utilities) and cyclone fired [which are important producers of electricity but which have essentially vanished from new construction (18)]. Dry-bottom boilers produce more fly ash in proportion to bottom ash than the other types. Pulverized-coal-fired boilers are also more adaptable to combustion modifications to reduce NO_x emissions.

BOILER DESIGN AND OPERATION

Modern dry-bottom boilers fired with pulverized coal generally have a generating capacity of a few hundred to several hundred megawatts (MW), with a representative average of 500 to 600 MW (18). The design of these boilers and

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other types is discussed in detail in manuals published by boiler manufacturers (19,20). In side view, the boiler somewhat resembles a compact inverted U, as shown in Figure 1. The furnace for a typical (500 to 600 MW) utility boiler is a rectangular vessel 40 to 50 feet on a side and up to 200 feet high (boilers designed to burn low-rank coals are usually larger for a given capacity than those designed to burn high-rank coals). At the top is a horizontal pass that contains superheater and reheater tubes and at the rear is a vertical pass that contains additional superheater and reheater tubes and an economizer that heats the boiler feedwater. The furnace and most or all of the horizontal and vertical passes are lined with tubing that serves as part of the boiler water-steam system. After passing through the economizer, the flue gas is passed through an air heater that heats the combustion air. The flow of combustion air is controlled by forced-draft (FD) fans and the flow of flue gas is controlled by induced-draft (ID) fans, generally situated downstream from the fly ash collection equipment if possible.

Air heaters recover heat from the flue gas by heating the combustion air. This also improves the combustion performance and provides the heated air necessary to dry the coal in the pulverizers. Air heaters can be either recuperative--which are essentially large shell-and-tube heat exchangers--or regenerative, in which a heat absorbing surface is exposed alternately to flue gas and combustion air. The Ljungstrom air heater is typical of regenerative air heaters (20). It consists of a large horizontal rotor with steel sheets to provide a large surface. The rotor turns slowly in a housing with elaborate seals to provide parallel paths for flue gas and combustion air so that sections are exposed alternately to flue gas and combustion air. Air heaters are exposed to flue gas near the acid dewpoint and cool combustion air and are subject to corrosion and plugging. Processes that affect the flue gas properties upstream from the air heater may make modifications to the air heater necessary.

The coal is reduced to a fine powder--typically about 100 micrometers in size--and dried in pulverizers--typically consisting of rollers that ride on a rotating bowl-shaped grinding table (21). A portion of the heated combustion air--called primary air--is passed through the pulverizer to dry the coal and transport it to the furnace burners. The burners, up to a few dozen in number, are arranged in various arrays in the front, front and back, or corners of the furnace. The coal-bearing primary air is injected through a central nozzle in the burner and all or some of the remaining combustion air, called secondary air, is injected through a larger orifice surrounding this nozzle. Burner design philosophy has undergone a revolution as the result of NO_x emission regulations stemming from the Clean Air Act Amendments of 1970 (Public Law 91-604). Before these regulations, burner design had concentrated on the production of a hot, compact flame; conditions that increased carbon burnout; improved flame stability; and minimized operating problems such as slagging, but which also increased the formation of NO_x. In the 1970s, the objective became the development of burners that decreased the maximum flame temperature and minimized the concentration of oxygen present at the higher temperatures.

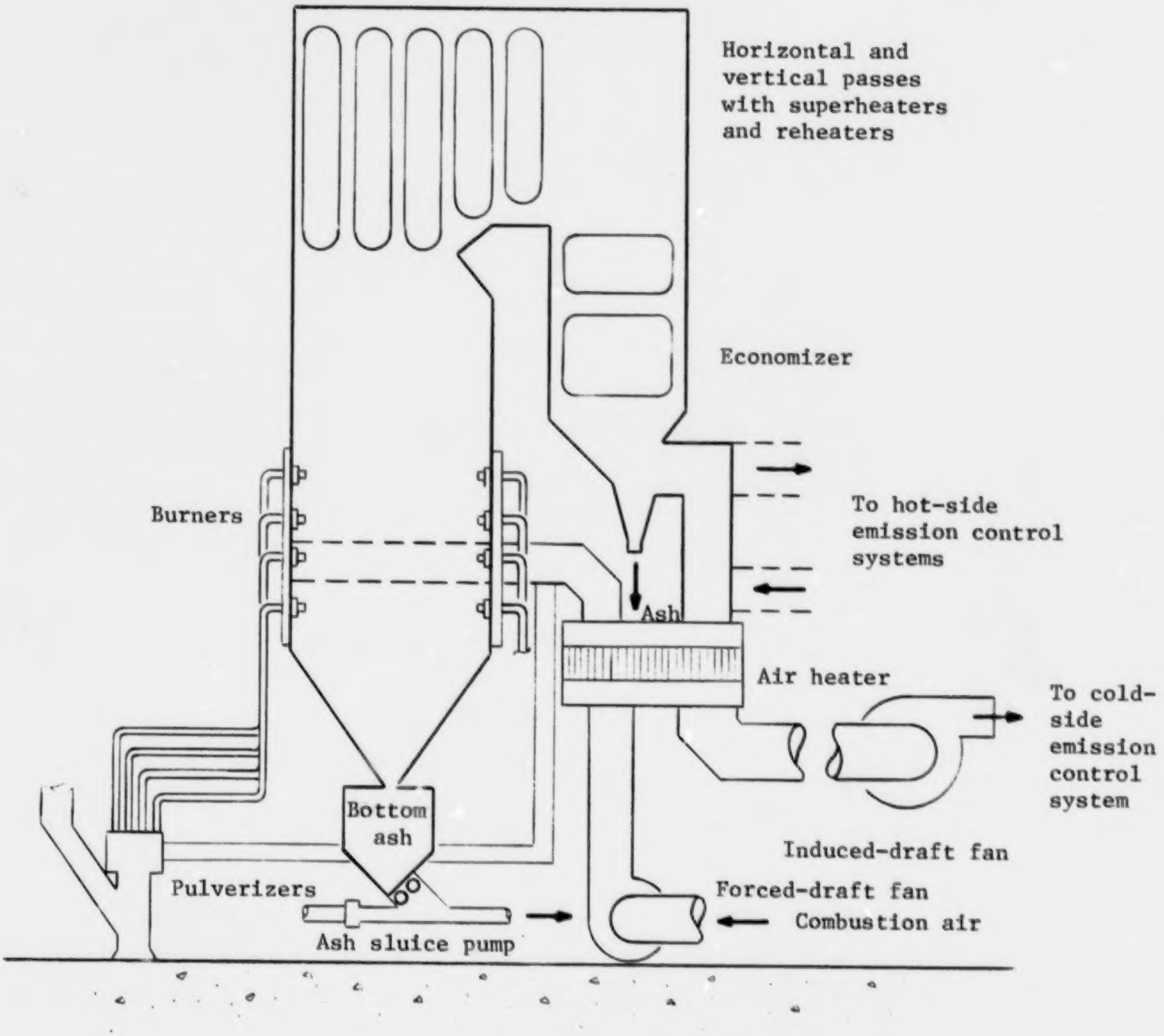
Horizontal and vertical passes with superheaters and reheaters
Economizer
Burners
To hot-side emission control systems
Ash
Air heater
To cold-side emission control system
Bottom ash
Induced-draft fan
Pulverizers
Forced-draft fan
Combustion air
Ash sluice pump

Figure 1. Typical pulverized-coal-fired boiler configuration.

The coal particles suspended in the combustion air burn in less than a second. A small part of the nitrogen in the coal and air is converted to NO_x during this time. Sulfur is oxidized to SO_2 and a small percentage is further oxidized to SO_3. The NO_x and the sulfur oxides, except for a small percentage trapped in the ash particles, remain as gases in flue gas. Some of the mineral matter in the coal such as carbonates and some minor elements are decomposed to gaseous oxides. Other mineral matter such as alkali metals and some trace elements vaporize and, subsequently, condense as submicrometer particles or on the surfaces of other particles. Most of the mineral matter is wholly or partly fused and solidifies as small particles that are typically spherical and porous or hollow. The larger particles and agglomerates of particles, along with slag dislodged from the furnace, fall through a throat in the bottom of the furnace as bottom ash; the rest, usually about 80% of the total, is carried out of the furnace as fly ash in the flue gas.

The flue gas typically enters the horizontal pass at about 2,000°F and the economizer at about 800°F. It usually enters the air heater at about 700°F where it is cooled to about 300°F. The temperature of the flue gas leaving the air heater is usually determined by the need to avoid the corrosive effects of sulfuric acid condensation. The acid dewpoint is usually in this range, depending on the sulfur content of the coal and the fraction converted to SO_3.

The water content of the flue gas under ideal conditions is determined by the inherent moisture (that contained in the coal as mined) and surface moisture contents of the coal, the water content of the combustion air, and the water formed during combustion. This usually produces a water dewpoint of roughly 120°F to 130°F in the flue gas, lower for bituminous coals with low inherent moistures and higher for subbituminous coals with high inherent moistures. Several factors can contribute to higher flue gas water contents, however, often in ways that produce wide unpredictable variations. Among these are variations in the surface moisture on the coal and humidity of the air caused by precipitation, sootblowing and sootblower malfunctions, and steam leaks. Normally the flue gas leaving the boiler is well above the water saturation temperature; the variations can, however, have important effects on the design concepts and operation of spray dryer FGD systems.

The bottom ash that falls through the throat at the bottom of the furnace is collected in a bottom ash hopper. Other systems are used but it is typically a water-filled multiple-vee-bottom hopper that is periodically emptied by sluicing the ash through clinker grinders into a pump that transports the ash as a water slurry to a pond or dewatering system.

Most of the fly ash remains entrained in the flue gas and is removed in the fly ash collection equipment--ESPs or baghouses. Some, however, settles in the boiler or post-boiler equipment where it is necessary to place hoppers and handling equipment to remove it. Usually hoppers are installed at the base of the vertical pass containing the economizer and on the air heater. SCR reactors have usually been equipped, as they are in this study (except for the case with a hot-side ESP preceding the SCR reactor), with a hopper bottom to collect ash.

Fly Ash

In appearance, fly ash is a light powdery material ranging in color from earthy brown to black with somewhat the texture of a gritty silt. The geometric mean diameter is usually between 10 and 20 micrometers, with 1% to 10% below 1 micrometer and about 90% below 100 micrometers (22), a size distribution that encompasses clay through fine sand. Most fly ash particles consist of vitreous spheres that are frequently hollow (23). Others consist of fragments of spheres, irregular porous fragments, agglomerates, and char. The major chemical constituents are silicon, aluminum, and iron, which occur primarily as a variety of vitreous and crystalline silicates and oxides (24). The calcium, magnesium, and sodium contents seldom exceed 2% each in eastern bituminous coals; in western subbituminous coals and lignites, however, the calcium content usually exceeds that of iron and is usually in the 10% to 20% range. The magnesium and sodium contents of western coals are also usually higher than those of eastern coals. The carbon contents, which depend on the boiler operating conditions, are often less than 1% but may temporarily exceed 20%.

Fly ash is rich in minor and trace elements, a result of the process of coal formation in which 25 to 40 elements are abnormally concentrated (25). Many of these elements, among them those with potential harmful effects such as antimony, selenium, arsenic, and lead, are concentrated in the fly ash portion of the ash during combustion. There is also a variation of chemical composition with particle size and, in some cases, between the surface and interior of the ash particles.

Bottom Ash

Bottom ash is a relatively innocuous material similar in physical properties to a sandy gravel (26). Most of the particles range from 0.2 to 10 mm (less than 1/100 to about 3/8 inch) in size and range in texture from dense pieces of slag through rounded and angular vesicular particles to porous sintered aggregates. It has the same bulk chemical composition as fly ash but is depleted in the more volatile elements and is less reactive.

ASH HANDLING

The ash-handling systems consist of the hoppers associated with the boiler and emission control equipment, hydraulic and pneumatic transporting systems, and dewatering and storage equipment. Several methods and numerous variations of methods adapted to particular requirements are used. There has been, however, a trend toward methods that reduce or eliminate the use of water sluicing, the traditional method of transporting ash to the disposal site. There has also been at least the beginnings of a trend to incorporate improved technology in the form of improved hopper designs, transporting systems, and continuous ash removal systems into utility ash-handling systems. These and conventional ash-handling systems are discussed in detail in manufacturers' publications (20,27).

The equipment associated with the slurry-handling and transport systems is subject to severely abrasive conditions. Hard steels and iron are used for pumps and lines, often with wear plates or ceramic inserts at points of extreme wear such as nozzles and bends. Sometimes pipes lined with abrasion-resistant materials such as basalt are used. Corrosion and scaling can also be a problem in closed-loop water systems, necessitating water softening treatment and pH control.

Bottom Ash

Bottom ash hoppers are equipped with clinker grinders (two opposed steel-toothed rollers) through which the ash is periodically sluiced into a transportation pump by jets of water. The pump may be either a water ejector or a centrifugal pump. Water ejectors are simpler, less prone to plugging and air locks but centrifugal pumps produce higher heads and can be staged to provide higher pressures if necessary. The ash is transported as a 1% to 6% solid slurry at velocities up to about 700 ft/min, either to a temporary or permanent disposal pond or to a dewatering system.

Bottom ash dewatering systems (27) consist of dewatering bins into which the ash is pumped from the boiler hopper. Water drains from the ash into a settling tank to remove fines and then drains to a storage tank for treatment and reuse. The ash in the dewatering bin is dumped to a truck for removal to the storage or disposal area. Often the water must be treated to control scaling and adjust extreme pH levels.

A different type of bottom ash system called a submerged scraper conveyor or submerged drag bar chain conveyor, widely used in Europe, is now offered by several U.S. vendors and is coming into use in the United States (20). The boiler ash hopper is a water-filled low-profile hopper containing a continuous drag bar conveyor in the bottom. The drag bar conveyor (essentially two chains at each side of the hopper bearing transverse bars) operates continuously, drawing the ash to the end of the hopper and up an inclined dewatering screen. The ash can be passed through a clinker grinder and removed by conveyor or by trucking.

Fly Ash

Fly ash removal is frequently complicated by difficulties in removing the ash from the collection hoppers. Fly ash is usually somewhat hygroscopic and may pack and cake if it is allowed to approach the acid or water dewpoint temperature. Western coals also frequently produce ash with cementitious properties that add to the caking problems. The hoppers must be carefully designed and the system carefully operated to ensure efficient operation. The hopper bottoms generally slope at 55 degrees or more from the horizontal to a center outlet a foot or more across. The hoppers must also be heavily insulated and may be equipped with heaters, vibrators, and heated-air fluidizers. Economizer ash hoppers may be designed for continuous removal, for collection of the ash in water, or for isolation of the ash from the hot flue gas because of the tendency of economizer ash to form agglomerates--or to burn if it is rich in carbon.

Fly ash is usually removed from the hoppers intermittently through air lock valves by a vacuum or pressure pneumatic system. Usually the system is cycled so that one or a group of hoppers are emptied at a time. Vacuum systems allow the use of simple air lock valves and supplement the action of gravity with suction to remove ash from the hoppers. They are, however, limited in the distance that the ash can be transported and they are less efficient at higher altitudes. Pressure systems (which usually operate below 15 psig) require more complex air lock valves but they have higher capacities. Vacuum systems and most pressure systems operate in an overall "dilute phase" in which the ratio of fly ash to air is 20 to 1 or less and the velocities are 1,500 ft/min or more. "Dense phase" pressure systems with much higher ash-to-air ratios and lower velocities are coming into use, however (28).

In large complicated hopper systems, such as those associated with large baghouses, a vacuum-pressure system may be used. This allows the use of simplified air lock valve systems. The ash is conveyed by the vacuum system to a nearby separator mounted on a surge tank, from which it is conveyed to storage silos by the pressure system.

The vacuum in vacuum systems can be supplied by a mechanical pump or by steam or water eductors. If water eductors are used, the ash can be drawn directly into the eductor and mixed with the water. The slurry of a few percent solids is discharged to a deaerator tank and flows by gravity to a disposal pond. The ash can also be removed in cyclone-baghouse collectors and stored in a surge tank, from which it is removed, slurried, and pumped to a pond. Since the fly ash cannot be readily dewatered, once slurried for transportation by sluicing, some form of temporary or permanent pond disposal is inevitable. Because of the practical and environmental problems associated with ponding, this once almost universal method of fly ash disposal is giving way to dry collection and disposal methods. Increasing use of fly ash has also encouraged dry collection methods.

In dry disposal, the ash is removed from the conveying air using the same cyclone-baghouse collection system and stored in metal or concrete silos. The silos are usually elevated to permit direct loading into trucks or railcars, often through a moisture-mixer to reduce dusting problems.

CONTROL OF NITROGEN OXIDE EMISSIONS

The formation of nitrogen oxides is not an invariant function of the fuel characteristics but depends in part on the nature of the combustion process. NO_x (consisting of about 90% to 95% NO and 10% to 5% NO_2) is formed by the temperature-promoted reaction of oxygen with nitrogen in the combustion air (called thermal NO_x) and nitrogen in the fuel (fuel NO_x). The quantity formed depends on the time-temperature relationship of the combustion process, the amount of oxygen available at the higher flame temperatures, the quantity of fuel nitrogen in the coal, and the nature of the release of fuel nitrogen during combustion (29). Consequently, the emission of NO_x can be controlled to some extent by modifying the combustion process. This approach to NO_x emission control--"combustion modifications"--has proven widely successful in meeting many NO_x emission control requirements. Flue gas treatment, however, is technically capable of providing larger reductions in NO_x emissions and is thus a potential alternative means of attaining low emission levels.

Combustion modification techniques consist of physical modifications to the fuel burners (this is most feasible in pulverized-coal-fired boilers) and the furnace to reduce the flame temperatures and the duration of higher temperatures and to limit the amount of oxygen available during the higher temperatures. In general, the burners are designed to produce a low-turbulence flame with "staged combustion" to maintain an oxygen-deficient atmosphere during the combustion phases most favorable for NO_x formation. To reduce the availability of oxygen, the lowest practical excess air level is used and some of the combustion air may be admitted through air ports around or above (overfired air) the burners, or burners that are not supplied with coal may be used to admit air ("burners out of service"). The area of the furnace occupied by the burners may also be increased to reduce the flame intensity and the heat-absorbing area of the furnace may be increased to induce a rapid initial cooling of the combustion products.

Several successful low-NO_x burners and furnace designs are now offered commercially (30). They have become standard equipment, along with the associated furnace modifications, for new construction to meet most NO_x emission control requirements. Such burners combined with the injection of pulverized limestone for SO_2 control--for example, the limestone injection multiple burner or LIMB process--are also being developed (31). Combustion modifications are not, however, without problems associated with an oxygen-deficient atmosphere such as slagging and tube corrosion. The present NO_x emission limits are, in part, a balance between the possible reductions attainable and the severity of operating problems associated with combustion modification techniques (32).

Several general types of flue gas treatment processes to control NO_x emissions have advanced to various stages of development and use, including wet-scrubbing and dry adsorption processes that frequently also incorporate SO_2 removal. Wet processes have been essentially abandoned because of their complexity and expense (33); the development of dry adsorption processes is continuing, however (34). The most highly developed processes involve injection of NH_3 into the flue gas to reduce the NO_x to molecular nitrogen. These appear to be the most economical methods of flue gas treatment and do not produce a waste product. Processes of this type are in commercial use in Japan, although limited information is available on their use on coal-fired boilers (35). They are called selective reduction processes because the reaction of NH_3 with NO_x rather than other flue gas components is favored (other troublesome reactions such as the formation of ammonium-sulfur compounds occur, however). Adequate uncatalyzed reaction rates occur in a narrow temperature range at about 1,700°F to 1,800°F which has led to the development of processes in which the NH_3 is injected into the furnace above the burner zone. These processes--called selective noncatalytic processes (SNR)--are actively being developed (36). Most of the processes, including those in commerical use, involve a catalyzed reaction at a lower temperature and are called SCR processes. The catalyzed reaction occurs in a temperature range between about 600°F to 750°F, allowing treatment of the flue gas after it has passed through all of the boiler components except the air heater. The flue gas from the boiler economizer is treated with NH_3 and ducted to a reactor containing beds of catalyst and is returned to the boiler air heater. Usually it is necessary to modify the air heater because of reaction byproducts and unreacted NH_3 in the treated gas.

Selective Catalytic Reduction

A typical SCR system consists of one or more trains of vertical reactor vessels connected by ducts to the economizer outlet. The ammonia is injected into this duct as a dilute mixture in air. The flue gas containing the ammonia flows downward through the reactor, passing through multiple layers of catalyst, which is in the form of honeycomb-like blocks or bundles of tubes. The flue gas is ducted from the reactor to the air heater. If not preceded by an ESP, the reactor has a hopper bottom to remove fly ash that settles in the reactor. An economizer bypass to supply hot flue gas to the reactor inlet is often provided to maintain the necessary reactor temperature at low loads.

The flue gas flow rate and temperature, the reactor pressure drop, the inlet NO_x and oxygen contents, and the outlet NO_x and ammonia contents are monitored to serve as a basis of control for the economizer bypass, ammonia injection, and sootblower operations. Based on experience with commercial operations in Japan, there are unresolved problems in process control because there are no fully reliable methods of continuously monitoring low levels of ammonia in flue gas that contain NO_x, SO_2, and fly ash. A chemiluminescence method in which the ammonia is reduced or oxidized to nitrogen or NO and determined by difference is most commonly used. An ultraviolet absorption method has also been evaluated. All methods used so far, however, have suffered from interference from other components and incomplete conversion of ammonia (37).

The NH_3 is stored as an anhydrous liquid. It is vaporized and diluted with a gas--usually air--and injected into the flue gas upstream from the reactor through an injection grid that consists of an array of pipes with nozzles across the cross section of the duct. The dilution reduces the possibility of explosions and increases the volume of the injected gas to improve mixing with the flue gas. A mixing grid, downstream from the injection grid, consisting of a matrix of pipes, ensures complete mixing.

Chemical Reactions--

The catalyst promotes the direct reaction of NH_3 with NO_x to form molecular nitrogen and water.

$$4NH_3 + 6NO \rightarrow 5N_2 + 6H_2O \quad (1)$$

$$8NH_3 + 6NO_2 \rightarrow 7N_2 + 12H_2O \quad (2)$$

If oxygen is present, however, it also enters into the reaction.

$$4NH_3 + 4NO + O_2 \rightarrow 4N_2 + 6H_2O \quad (3)$$

$$4NH_3 + 2NO_2 + O_2 \rightarrow 3N_2 + 6H_2O \quad (4)$$

About 1% oxygen in the flue gas is sufficient to favor the reactions shown in reactions 3 and 4, which are also favored by higher temperatures. From a practical point of view, since flue gas normally contains at least several percent oxygen and since NO constitutes 90% or more of the NO_x, reaction 3 is the most significant. However, there has been limited data from commercial operations in Japan indicating the importance of reaction 1 occurring, in addition to reaction 3, since 80% NO_x reduction has been obtained at one commercial facility with an NH_3:NO_x mol ratio slightly less than 0.8 (37).

NH_3 also reacts directly with oxygen at a rate increasing with temperature.

$$4NH_3 + 3O_2 \rightarrow 2N_2 + 6H_2O \quad (5)$$

$$4NH_3 + 7O_2 \rightarrow 4NO_2 + 6H_2O \quad (6)$$

The catalyst also promotes the oxidation of SO_2 to SO_3 (based on vendor information, the percent SO_2 oxidized to SO_3 ranges from 0.5% to 2.0%) which, added to the few percent SO_2 normally oxidized in the boiler, increases the acid dewpoint of the flue gas and promotes formation of ammonium sulfate and bisulfate (38,39,40).

$$2NH_3 + SO_3 + H_2O \rightarrow (NH_4)_2SO_4 \quad (7)$$

$$NH_3 + SO_3 + H_2O \rightarrow NH_4HSO_4 \quad (8)$$

These salts exist as solids at temperatures of about 350°F or higher, depending on the concentrations of NH_3 and SO_3. The NH_4HSO_3 can form in the reactor at temperatures up to about 600°F where it forms a coating on the catalyst. To avoid this, vendors recommend a minimum operating temperature somewhat above 600°F to be maintained by hot flue gas bypass under low-load conditions (37). Formation of these salts in downstream equipment as the flue gas is cooled to 350°F or below is essentially unavoidable, though it can be reduced by minimizing the formation of SO_3 and the amount of unreacted NH_3--the NH_3 "slip" or "breakthrough"--in the flue gas. Ammonium sulfate salts typically form in the air heater as sticky corrosive deposits that usually necessitate modification of the air heater to reduce corrosion and plugging (41,42).

In addition to the certain effects of ammonium sulfates/bisulfates upon air heaters, there are potential effects of unreacted ammonia and ammonia salts upon downstream equipment such as ESPs, baghouses, FGD processes, and waste disposal which are not well defined and require further study (43,44). For example, ammonia may improve ESP collection performance but aggravates plate cleaning and discharging of ash from hoppers. Ammonium salts may blind the filter media in baghouses, requiring more frequent bag cleaning and bag replacement. Ammonia slip may benefit FGD systems by increasing SO_2 removal and reducing absorbent stoichiometry.

In addition to the effects above, there is concern for the effect of ammonia and ammonium salts on utilization of fly ash. Fly ash from the Shimonoseki unit in Japan, where ammonia slip has been maintained at very low levels, however, has been used in cement with no quality problems (37).

Catalyst--

The catalyst itself is actually a surface coating on--or component of--rigid metal or ceramic boxwork structures that are usually manufactured in elements of a standard size, usually a fraction of a meter in width and length and a meter in depth (the direction of flow). These elements, which are themselves called the catalyst, are assembled in metal frames called modules that are in turn placed on supports in the reactor to provide the desired surface area of catalyst. (Somewhat confusingly, the quantity of catalyst is frequently reported in terms of "space velocity," which is the volume of flue gas divided by the volume of catalyst. This relates directly to an area only for a particular catalyst design.) In some cases, the modules may be formed of individual pipe-like elements. The number of layers of modules is determined by the surface area of catalyst needed. The reactor is designed with removable sections, framework, and hoisting equipment to facilitate replacement of the modules. Sootblowers are also installed above and below each layer to clean the catalyst.

The catalyst is one of the most important features of SCR processes, as well as the primary distinguishing characteristic of the processes offered by different manufacturers. The initial catalyst cost is an important element of capital investment and catalyst replacement costs are--or are believed to be--by far the largest operating cost (45). The manufacturers have pursued active catalyst development programs that have produced unique physical configurations and, presumably, compositions. These designs are closely guarded proprietary information, which makes quantitative comparisons difficult, however.

Application of SCR processes to coal-fired boilers places severe demands on the catalyst because of the fly ash (which both erodes and coats the catalyst and can plug the reactor), the higher levels of flue gas components that can cause the catalyst to deteriorate, and the generally higher levels of SO_2. SCR processes were developed for clean flue gas using beds of catalyst balls. The rigid honeycomb catalyst was developed for dirty flue gas to reduce erosion and plugging and to permit cleaning with sootblowers. The catalyst modules are designed to provide a flue gas velocity that minimizes the pressure drop and erosion but which is high enough to prevent the fly ash from coating or plugging the catalyst. The dimensions of the passages in the catalyst (the pitch) range from about 5 to 20 mm, depending on the fly ash content of the flue gas and the design philosophies of the manufacturers.

The active components of the catalysts are usually based on vanadium and titanium oxides (46) but the actual compositions are proprietary. The catalyst can either be applied as a surface coating or may be incorporated into the support material, based on the assumption that erosion exposes fresh catalyst (47). In addition to erosion, the catalyst effectiveness deteriorates because of poisoning by alkali metals and some heavy metals, prolonged overtemperatures, and blinding or chemical masking, which can be wholly or partially reversible by sootblowing operations or washing. Inevitably, however, the catalyst must eventually be replaced. Vendors generally guarantee a 1-year catalyst life and lives of two years or more have been attained in some coal-fired applications in Japan (37) (the economic evaluations showing the large catalyst replacement cost are normally based on a 1-year life). It is uncertain whether the entire catalyst charge deteriorates uniformly or whether a partial replacement is satisfactory (48).

Some of the major objectives of the ongoing catalyst development efforts are to increase the resistance of the catalyst to fly ash abrasion, to reduce chemical deterioration caused by flue gas components such as alkali metals, to (improve the specificity to) reduce the formation of undesirable side products such as SO_3, and to reduce the pressure drop and tendency of fly ash to blind the surface or physically to plug the catalyst.

Another potential environmental effect of SCR applications is catalyst disposal. At the time of this study, both catalyst disposal and reclamation procedures were under review by the process vendors (38,39,40). Although most vendors indicated willingness to participate in spent catalyst disposal studies, no definite procedures have been established, primarily because none of the commercial SCR units have required a catalyst change (37).

Commercial Systems in Japan--

The commercial SCR systems operating (or soon to be operating) on coal-fired boilers in Japan are listed in Table 1. The Shimonoseki and Shin-Ube systems treat flue gas with a full fly ash loading. The Tomato-Atsuma and Takehara systems treat flue gas from which the fly ash is removed by hot-side ESPs upstream from the SCR system. All of the other systems treat flue gas from which the fly ash is removed downstream from the SCR system.

TABLE 1. SCR UNITS FOR COAL-FIRED UTILITY BOILERS IN JAPAN

Owner	Plant site	Boiler No.	 MW	 N/P[a]	Constructor	Completion
EPDC	Takehara	1	250	R	BHK[b], KHI[c],	1980
		2	700	N	BHK	1983
Chugoku	Shimonoseki	1	175	R	MHI[d]	1980
Electric	Shin-Ube	1	75	R	MHI	1982
		2	75	R	MHI	1982
		3	156	R	MHI	1982
	Mizushima	1	125	R	BHK	1983
		2	156	R	BHK	1983
Hokkaido Electric	Tomato-Atsuma	1	350[e]	N	BHK	1980
Kyushu	Omura	2	156	R	MHI	1982
Electric	Minato	1	156	R	MHI	1983
Joban	Nakoso	8	600	N	MHI	1983
Joint		9	600	N	IHI[f]	1983
Tohoku Electric	Sendai	2	175	R	BHK	1983
Tokyo	Yokosuka[g]	1	265	R	MHI	1984
Electric		2	265	R	MHI	1984

Source: Reference 37

a. New or retrofit.
b. Babcock Hitachi K.K.
c. Kawasaki Heavy Industries.
d. Mitsubishi Heavy Industries.
e. Subjecting one-fourth of the gas to SCR.
f. Ishikawajima-Harima Heavy Industries.
g. Currently uses oil and will use coal-oil mixture from 1984.

The Shimonoseki system is designed for 50% NO_x reduction in a flue gas containing about 55 ppm NO_x. The system has operated without trouble and has produced NO_x reductions of 50% and 55% at NH_3:NO_x mol ratios of 0.51:1.0 and 0.56:1.0. The system was designed for an ammonia breakthrough of 10 ppm or less; during the initial operation, the breakthrough was 1 ppm or less but later increased to 2 to 3 ppm without affecting the NO_x reduction. The catalyst has shown little change during the first two years of operation. It was initially expected to have a life of three or more years.

The Takehara system consists of two units in parallel on a 250-MW boiler. The system was designed for 80% NO_x reduction in flue gas containing 400 ppm NO_x with an ammonia breakthrough below 5 ppm. The system was started up in April of 1981 for tests and produced the 80% NO_x reduction at an NH_3:NO_x mol ratio of slightly less than 0.8:1.0. The system was placed in commercial operation in August 1982.

The Tomato-Atsuma system was designed for 90% NO_x reduction but has been operated to obtain 80% reduction. With NH_3:NO_x mol ratios of 0.85 and 0.95, NO_x reduction efficiencies of 81% and 92% were obtained, with ammonia breakthroughs of less than 2 ppm. The system treats one-fourth of the flue gas, which contains 200 ppm NO_x. The gas is combined with the remaining flue gas to produce an ammonia concentration of less than 1 ppm. The SCR system was the first designed by Babcock Hitachi for a coal-fired boiler and was designed with sufficient safety factors to ensure that it would not interfere with the operation of the boiler. The catalyst effectiveness had not changed after one year of operation.

The Shin-Ube system was started up in the spring of 1982 as part of a combined NO_x-SO_2 control system. The system is operated to obtain a 65% NO_x reduction, treating flue gas containing 400 ppm NO_x. It has a honeycomb catalyst with a space velocity of 4,000 hr^{-1} and an NH_3:NO_x mol ratio of 0.66. The ammonia breakthrough has been less than 1 ppm. The catalyst has a pitch of 7 mm, which is more efficient than the 10-mm pitch used at the Shimonoseki system.

ELECTROSTATIC PRECIPITATORS

ESPs consist of arrays of vertical collection plates several inches apart, interspaced with wire-like electrodes in a housing with a hopper bottom. The flue gas flows through the array parallel to the plates at velocities of a few feet per second, normally at a pressure drop of 1 to 2 inches H_2O. A voltage sufficient to create a corona discharge is placed on the electrodes, which, by a complex ionization process (16,49), produces a flow of charged particles to the grounded collection plates. Particles in the flue gas acquire a similar charge and migrate to the collection plates and adhere to them. The accumulating layer of particles is periodically dislodged by rapping the plates and falls into the collection hoppers at the bottom of the housing. The size of ESPs is expressed in terms of plate area per unit of flue gas volume--called the specific collection area (SCA)--in $ft^2/1,000$

aft^3/min. For normal applications in the past, SCAs of 300 to 500 have been typical of high-efficiency ESPs.

Since they were first used in the 1920s, ESPs have proven to be a highly satisfactory means of meeting most of the particulate emission control requirements of utilities (50). The removal efficiencies imposed by the particulate emission and opacity regulations of the past decade and the problems associated with some coals coming into increasing use have, however, placed rigorous demands on ESP technology (51). Emission regulations such as the 1979 NSPS require removal efficiencies well in excess of 99%; plume opacity restrictions have increased the importance of submicrometer particle collection, for which ESPs are less effective than for larger particles. At the same time, increasing use of low-sulfur coal and western coal often creates conditions that make the collection of fly ash in ESPs more difficult.

It is generally most desirable, operationally and economically, to locate ESPs downstream from the boiler air heater where the flue gas is coolest and contains no recoverable heat: the ESP can be smaller and does not have to be designed for the rigors of high operating temperatures and large temperature changes, no useful heat is lost, and the boiler design and ducting are unaffected. In some cases, however, these "cold-side" ESPs are impractical because of the ash electrical resistivities in this temperature range. In these cases, a "hot-side" ESP, which treats the flue gas from the boiler economizer before it passes through the air heater, may be more practical. Although more expensive and not without difficulties, hot-side ESPs have been used with increasing frequency to collect high-resistivity ash (52).

The effectiveness with which an ESP collects particles is determined by the potential of the electrodes, which determines the current density and field strength between the electrode corona and the plates and the potential on the charged particles. There are, however, factors that limit the maximum practical potential, among which the properties of the particles themselves and the temperature and composition of the flue gas are pivotal. All of these affect the electrical resistivity of the particles, which determines the rate at which the charge on the particles dissipates. Resistivities of about 1×10^7 ohm-cm are regarded as most suitable for collection in an ESP (50). Resistivities regarded as high may range upward to about 10^{13} ohm-cm. If the resistivity is too low--which is rare in utility applications--the particles on the plate lose their charge and become reentrained. More common, and encountered with increasing frequency, are situations in which the resistivity is too high. In this case, the ash adheres so firmly to the plates that it cannot be effectively dislodged. High resistivity also creates a high potential across the layer of particles on the collection plates that induces an electrical breakdown called back corona. The potential across the layer exceeds the dielectric strength of the gases in the layer, causing them to ionize and essentially destroy the capacity of the affected plate area to collect particles (53). To avoid these problems, it is necessary to operate at a lower voltage--which requires a larger SCA to attain the desired removal efficiency--or to operate at conditions that produce a lower resistivity.

The resistivity of the ash is a function of the mobility of the charge on the particle. At low temperatures, it is a surface phenomenon in which the charge is conducted by a surface film formed on the ash particles by condensed or adsorbed gases. As the temperature increases, the effectiveness of this mechanism decreases but conduction through the particle itself becomes increasingly effective. As a result, ash resistivities usually increase rapidly with temperature as surface resistivity increases, reaching a maximum between 250°F to 350°F, then rapidly decrease as volume resistivity decreases. Plotted with temperature as the abscissa, log resistivities resemble an inverted U with a peak near 300°F.

Sulfur trioxide is effective in forming a conducting film on the fly ash particles in the temperature range at which surface conductance is an effective charge transfer mechanism. Usually coals with moderate to high sulfur contents--over about 2% (50)--produce flue gas with sufficient SO_3 to allow efficient collection of fly ash in a cold-side ESP. Alkali metals, particularly sodium, are believed to be the primary charge carriers in volume conductance (54) and probably surface conductance (55). Other components of the ash are also believed to affect conductance: calcium, for example, increases the resistance and iron decreases it (56).

Eastern bituminous coals typically have a high-sulfur content, a low calcium-to-iron ratio, and a moderate sodium content (57). They produce a flue gas and fly ash that usually allow effective fly ash collection with an ESP at temperatures at which either surface or volume resistivity is the controlling factor. Western coals typically have a low-sulfur content, a high calcium-to-iron ratio, and a sodium content that may be a small fraction of a percent to several percent, depending on the coal (58). Generally, the fly ash from western coals has a high surface resistivity because of the low SO_3 content of the flue gas. The volume resistivity may be high or low, depending on the sodium content of the ash. In most cases, however, the volume resistivity is suitable for collection in an ESP operating at 600°F to 800°F.

In the United States, the evolution of ESP technology was largely based on the collection of ash from eastern high-sulfur bituminous coals which produces ash and flue gas conditions particularly favorable for collection in an ESP. In addition, removal requirements were seldom as stringent as many now required. These conditions were readily met by cold-side ESPs situated after the boiler air heater where the flue gas temperature is about 300°F. They economically and reliably provided removal efficiencies in the high 90% range. In a highly competitive market, design emphasis was placed on reduction of capital costs by minimizing the size and structural complexity of the ESP, a practice that came to be called "American design" philosophy (50). In contrast, ESPs constructed in other countries, often to collect ash less easily collected in an ESP than ash from high-sulfur U.S. coals, followed a "European design" philosophy that placed greater emphasis on efficiency, reliability, and structural competence.

When low-sulfur coals came into wide use in the 1960s, the performance of these cold-side American design ESPs sometimes proved disappointing because of the high resistivity of the ash, which reduced the collection efficiency and required rapping so vigorous that the ESPs were physically damaged. To improve the efficiency, hot-side ESPs, several hundred of which are now in use by utilities (16), were developed. Hot-side ESPs treat flue gas from the boiler economizer before it passes through the air heater where the higher 600°F to 800°F temperature provides a low volume resistivity independent of the flue gas SO_3 content.

Unexpected problems have also occurred with hot-side ESPs, however. Some have been mechanical problems caused by the higher operating temperatures and larger temperature changes but high resistivities have also been encountered at the higher flue gas temperatures (59). Usually these are associated with low-sodium coals in which there are insufficient alkali metal ions to serve as charge carriers. There have also been instances in which the performance of the ESPs deteriorated with time. This is believed to be caused by a gradual depletion of sodium in an ash layer on the collection plates that is not dislodged when the plates are rapped. The permanent layer gradually becomes depleted in sodium, causing an increase in the apparent resistivity of the ash (60).

Very high removal efficiencies, the continuous nature of many emission regulations, and the more severe operating conditions also place severe demands on the design and operation of ESPs (61). The American design philosophy was criticized (62) because it sometimes did not provide the necessary reliability and continued high level of efficiency to meet these requirements. Gas flow characteristics also became more important since it became necessary to reduce the amount of flue gas that bypasses the collection area (sneakage) and ash that evades collection or becomes reentrained because of velocity variations and turbulence (63).

Flue gas conditioners have frequently been used to improve the performance of ESPs (50). Usually this has been done to increase the efficiency of an existing ESP that did not meet the design performance or to increase the efficiency to meet new emission regulations; conditioners have seldom been a factor in design considerations (61). For cold-side ESPs, SO_3 is a common conditioner, although ammonia and a number of similar substances have been used. Numerous proprietary materials are also available, although many do not affect the ash resistivity itself but alter the physical characteristics of the ash to reduce disintegration of the ash layer during rapping or to make it easier to dislodge (64). Conditioners can also be used with hot-side ESPs. In this case, however, a sodium compound is injected into the boiler or flue gas to increase the volume conductance of the ash (65).

The problems encountered with ESPs required to meet strict emission limits, particularly with low-sulfur coals, have created a pessimistic outlook for the future use of ESPs in utility applications (51). The difficulties have evoked a spate of research and development efforts in ESP technology, a field formerly left largely to the manufacturers of ESPs. EPA, the Electric Power Research Institute (EPRI), and others have sponsored symposiums intended to disseminate information on ESP technology (66) and both, along with private organizations, are actively conducting studies of ESP technology. These studies involve both the improvement of existing ESP designs and the development of advanced designs. Among the latter are pulse energization and 2-stage ESPs. Pulse energization, which is commercially available technology (67), consists of applying a high-frequency voltage pulse to the energizing voltage, which increases the efficiency of the ESP without corresponding increases in sparking and back corona. In 2-stage ESPs, the particle charging zone is separate from the collection zone (68).

FABRIC FILTERS

The fabric filters used by utilities usually consist of bags about 1 foot in diameter and 30 to 40 feet long suspended vertically by the closed end in compartments of 100 to 300 bags. The open, bottom ends are attached to a tube sheet, also called a thimble sheet, that divides the compartment into an upper section containing the bags and a lower section with a hopper bottom. Flue gas is admitted to the lower section and flows upward through the bags into the upper section, from which it is ducted to the stack. A baghouse for a large utility boiler typically consists of up to several dozen compartments to provide an "air-to-cloth" ratio (aft^3/min of flue gas per ft^2 of filter area) of about 2, with provisions for compartments that are out of service for cleaning and maintenance (69). The material and fabric construction is a subject of great importance because of its bearing on bag durability. Several temperature-resistant materials and numerous fabric constructions are potentially useful but the technology in this area is still evolving (70). In commercial installations, woven glass fabrics with a Teflon or silicon coating, are usually used.

To clean the bags, each compartment has provisions for circulating reverse air: The flue gas is diverted to other compartments and clean air or gas is circulated through the bags in the opposite direction of the flue gas flow. The bags collapse, dislodging the accumulated cake, which falls into the hopper in the bottom of the compartment. A bag shaker is sometimes used to supplement the reverse airflow and in a few cases, shakers alone are used.

This type of baghouse is referred to as a "conventional" or "low-ratio" (air-to-cloth) design. A few utilities have installed high-ratio designs (71) in which the fly ash is collected on the outside of the bags. These designs have air-to-cloth ratios of 6 or more and about the same pressure drop as low-ratio designs. The bags, usually felted instead of woven, are suspended from a tube sheet at the top of the baghouse compartment and are supported internally by a metal frame or cage. The bags are cleaned while in service by directing a pulse of high-pressure air into the open top of the bags which causes a bulge to travel down the length of the bag.

In comparison with ESPs, baghouses have higher pressure drops--design pressure drops are usually about 6 inches H_2O compared with about 2 to 3 inches H_2O for an ESP--an important consideration since each 1-inch H_2O increase in pressure drop is estimated to cost $2 million in operating costs over the life of a 1,000-MW plant (72). Baghouses are, however, relatively unaffected by changes in flue gas volume and by ash and flue gas properties such as those caused by load changes and coal variations that affect ESP removal efficiencies, and they are somewhat better collectors of the submicrometer particles that cause opacity problems and which are attracting attention because of their physiological effects (73).

Fabric filters were first investigated for utility use in the 1960s in conjunction with injection of dry alkalis into the flue gas to remove SO_2 (74). A cautious interest in fabric filters for fly ash control by utilities developed in the 1970s, with the advent of stricter limitations on particulate emission and opacity, as scaled-up and redesigned ESPs to meet these limitations became much more expensive and sometimes failed to meet the design specifications (61). Recently, this interest has grown as generally favorable experience with baghouses has accumulated (75).

The reluctance to adopt baghouses was not due to doubts of their efficiency--fabric filters almost inevitably remove in excess of 99% of the particulate matter in flue gas passed through them--but of their operability and durability: the effect of wet particulates on the pressure drop when operating below the water dewpoint, corrosion when operating below the acid dewpoint, and particularly the effect of bag life on operating costs. Bag life is a critical factor in the practicality of baghouses. A subbituminous coal-fired, 500-MW power plant, for example, requires a baghouse containing in excess of 10,000 bags for fly ash control (about 2 million aft^3/min of flue gas and an air-to-fabric ratio of 2). A short bag life would have enormous effects on the operating costs of a baghouse.

The first baghouse installations at utility power plants were started up in 1973 at the Nulca Station of the Colorado Ute Electric Cooperative, and at the Sunbury Station of the Pennsylvania Power and Light Company (16). Neither were particularly typical of utility power plants--the Nulca installation consisted of three baghouses on three 12-MW stoker-fired boilers and the Sunbury installation of four baghouses on four 38-MW boilers that burned a blend of anthracite, bituminous coal, and petroleum coke--but the operation of the baghouses was followed with considerable interest (16,76).

Both installations provided reliable removal efficiency well in excess of 99% and an essentially clear stack (near 0% opacity) over a range of boiler operating conditions. Bag life at the Nulca Station was initially low but was improved by mechanical improvements. By 1976, several other utilities were operating or installing baghouses (77). By early 1983, 84 baghouses with a total capacity of about 11,000 MW were in operation on utility boilers and an additional 32 with a total capacity of 10,000 MW had been contracted, including some high-ratio, pulse-jet designs (69).

A considerable amount of published information on baghouses in utility use has accumulated. Both EPA and EPRI have published proceedings (78) of symposiums that dealt with utility fabric filter technology. EPRI has published an evaluation of a large baghouse in operation at the Kramer Station of the Nebraska Public Power District (79), economic studies (80), and evaluation of design aspects (72). EPRI also conducted evaluations at a 10-MW test facility at the Arapahoe Station of the Public Service Company of Colorado (81).

SPRAY DRYER FGD

Spray dryer FGD--also called dry scrubbing--appeared in the late 1970s and quickly gained acceptance among utilities for control of SO_2 emissions from power units burning low-sulfur coal. From 1977, when the first commercial utility system was contracted, through early 1983, 15 commercial utility systems were contracted and 3 full-scale demonstration systems were operated (82). During this period, 12 vendors developed and placed spray dryer FGD systems on the market (83). The total capacity of the commercial systems is about 6,200 MW, about 10% of the total operating and contracted FGD capacity in the United States.

The application of spray dryer techniques to FGD is a readily appreciable solution to many of the problems and economic penalties of wet scrubbing. In basic concept, it is identical to standard and widely used spray dryer technology (84). An atomized slurry or solution of absorbent is dispersed in the flue gas under conditions that result in evaporation of sufficient liquid to form solid particles while the droplets are suspended in the gas. SO_2 reacts with the alkali in the absorbent to form calcium sulfur salts, continuing to do so at a diminishing rate as long as the particles are in contact with the flue gas and residual moisture remains in the particles. Some or all of the particles, consisting of calcium salts and unreacted alkali, remain entrained in the gas and are collected in a fabric filter baghouse or ESP as a dry granular waste. The only liquid system involved is that needed to prepare and meter the absorbent liquid to the absorber. The many problems associated with the liquid systems in wet-scrubbing processes--corrosion, erosion, plugging, and scale formation--are reduced or eliminated, as are the costs for these systems. In addition, the flue gas is not cooled to the saturation temperature and reheating costs are reduced or eliminated. Since particulate collection is an intrinsic part of the process, fly ash collection can be combined with collection of the FGD waste, using a single particulate collection system and producing a single dry waste that can be disposed of in a landfill without further processing (85).

Balanced against these manifest advantages, however, are some limitations. The contact time of the flue gas with the liquid phase is a period of a few seconds and a reactive absorbent, often at high stoichiometries, is essential. As a result, absorbent costs are high, particularly for high-sulfur coal applications. This, perhaps more than intrinsic limitations on removal efficiencies, has focused interest in spray dryer FGD on low-sulfur coal applications.

The rapid ascension of spray dryer FGD to a prominent position in SO_2 emission control is due in part to the nature of the process--which is essentially a combination of spray drying and particulate collection technologies, both of which are mature and widely used industrial technologies. Adoption of these technologies to FGD, however, required the development of processes quite dissimilar to industrial applications in some respects, particularly in size and control requirements. Typically, but not exclusively, the vendors of spray dryer FGD systems are manufacturers of particulate collection equipment who developed the necessary spray dryer technology or formed accordances of various natures with spray dryer manufacturers. The rapidity with which spray dryer FGD evolved is illustrated by the fact that most vendors did not begin development work until the late 1970s (the first pilot plants were operated at power plant sites in 1977); in some cases, commercialization proceeded almost simultaneously with development.

The development of spray dryer FGD was also aided by the lengthy efforts to develop dry-injection FGD. A major impetus was the dry-injection pilot plant at the Basin Electric Power Cooperative's Leland Olds Station, which was used to develop an FGD system for the Coyote Station being built in North Dakota by a consortium of power companies (86). The most practical, if not the only practical, absorbent proved to be nahcolite, a natural sodium bicarbonate mineral, which appeared likely to be unavailable in the necessary quantities. Evaluations of other absorbents led, in 1977, to the use of a spray dryer to improve the reactivity of the absorbent. The Rockwell International Corporation (RI), which had been evaluating a spray dryer as part of a sulfur-producing recovery process, provided the spray dryer and the Wheelabrator-Frye, Inc., baghouse, previously used for the dry-injection tests, as the collection device. This led to a joint venture--since dissolved-- by RI and Wheelabrator-Frye to market spray dryer FGD systems and to a contract for a commercial system for unit 1 at the Coyote Station. Shortly thereafter, three other companies, all of whom are now spray dryer FGD vendors, operated pilot plants at Basin Electric or Otter Tail Power Company power plants--a joint venture of the Joy Manufacturing Company and Niro Atomizer, Inc. (Joy/Niro); the Babcock & Wilcox Company (B&W); and the Carborundum Company (now Carborundum Environmental Systems)--in competition for commercial systems at Basin Electric's Antelope Valley Station unit 1 and Laramie River Station unit 3. These tests resulted in contracts for three systems with a total capacity of 1,500 MW.

For processes that have gained such widespread commercial acceptance, there is relatively little comprehensive design information available, in part a result of their rapid development, which has allowed little time for such information to accumulate and, in part, a result of the lack of long-term institutionally sponsored studies with published results. Most of the pilot-scale studies have been conducted by the vendors who are dedicated to particular design concepts or who regard some of their results as proprietary information. Commercial or large demonstration systems have not yet provided comprehensive data; only the startup phase of the Coyote Station soda-ash system (87), the operation of demonstration systems at the Northern States Power Company's Riverside Station (88), and the City of Colorado Springs' Martin Drake Station (89) have been reported. EPA has sponsored pilot-plant studies and published summaries (90) of spray dryer FGD applications. EPRI has published a status review (83) and initiated a 1-year pilot-plant study in 1982 (91). The Department of Energy (DOE) also conducted a pilot-plant study (92).

Overall, however, comprehensive design and comparative performance data on spray dryer FGD are scarce and the relative merits, if any, of particular design philosophies adopted by vendors are difficult to assess. In most cases, only summary descriptions of the various vendors' designs have been published.

Although the main thrust of spray dryer FGD has been directed toward low-sulfur coal applications, there has been interest in high-sulfur coal applications. There have been tests in which satisfactory removal rates were attained with high-sulfur coals (88) and some vendors are willing to guarantee high removal rates at coal sulfur levels in the range of 3% to 4% (83). There is, however, a practical limit at which the quantity of absorbent liquid necessary to supply the absorbent required produces a saturated flue gas, a limit that may be reached at a coal sulfur content of about 4% (83).

The economics may also be a limiting factor because of the large quantities of absorbent required in high-sulfur coal applications. In a TVA economic study (15), for example, the absorbent costs for a lime spray dryer FGD system for a power unit fired with 0.7% sulfur lignite were $1.7 million per year, while the absorbent costs for a power unit with the same capacity fired with 3.5% coal were $8.4 million per year--50% of the total operating costs.

The spray dryer FGD process consists of the absorber (spray dryer); a baghouse or ESP to collect the solids, which also serves to collect the fly ash; and ancillary feed preparation and waste-handling systems. The absorbent liquid is introduced into the absorber through either a rotary atomizer or 2-fluid nozzle atomizer as very fine droplets that evaporate to solid particles in a few seconds at most (the design residence time in the spray dryer is 6 to 10 seconds). SO_2 is absorbed by the droplets and reacts with the absorbent to form the same sulfur salt wastes as are produced in wet scrubbing. The nominally dry particles, which consist of unreacted absorbent and sulfur salts, continue to react with SO_2, although at a much slower rate while they are in contact with the flue gas. If a baghouse is used, this contact is an important mechanism of SO_2 removal; from 5% to 20% of the total SO_2 removal is attributed to reactions that occur as the flue gas passes through the collected absorbent on the bags (83). For ESPs where there is less intimate contact between the waste and the flue gas, additional removal is significantly less. Because of this and other reasons (some vendors manufacture only baghouses), baghouses have been preferred over ESPs--only two of the commercial utility systems have ESPs.

Most vendor designs follow typical spray dryer configurations in which the gas flows downward through a cylindrical vessel with a conical bottom, with the atomizer or atomizers mounted at the top, and leaves through a duct in the bottom or side. The side configuration (with the opening centered in the vessel, facing down) permits collection of some of the particulate matter in the bottom of the absorber. The cylindrical design is the most practical when rotary atomizers are used and is also suited to 2-fluid nozzle atomizers. The 2-fluid nozzles are equally suited to other configurations, however. In contrast to the cylindrical design now favored by other vendors, the B&W design, which uses 2-fluid nozzles, is a horizontal vessel with a rectangular cross section.

Partial flue gas bypass may be used, either for reheat or as a safety factor to ensure dry operation of the particulate collection equipment. In some cases, "warm gas" from the air heater outlet is used; in other cases, "hot gas" from the economizer outlet is used. Vendors appear to differ on the value of flue gas bypass--about one-half of them provide it and the rest provide it "if necessary" (83).

The individual absorbers of the commercial systems range up to 192 MW in size, the largest being the horizontal rectangular chamber installed by B&W on unit 3 of the Basin Electric Company's Laramie River Station. The largest cylindrical absorber is 51 feet in diameter (158 MW) and the smallest is 46 feet in diameter (110 MW). The use of a single rotary atomizer places limits on the practical size of the absorbers because of the flow rates and spray pattern attainable. This has led to special gas flow arrangements in large single-atomizer designs and the unusual (for spray dryer technology) use of multiple rotary atomizers. The 2-fluid nozzles, which have low capacities, can be used in any number and arrangement and place no particular restrictions on the size or shape of the absorber. This is discussed more fully below.

It is essential to the effective operation of a spray dryer that the absorbent liquid be broken into very fine droplets--probably no larger than a few hundred micrometers in size--and uniformly dispersed in the flue gas. Either rotary or 2-fluid nozzle atomizers, both standard equipment in conventional spray drying, are used. Each has arguable advantages and disadvantages (83) and it is possible that either may be most suitable, depending on the particular system (93). Thus far, vendors have favored rotary atomizers: seven use rotary atomizers and five use nozzles, though one vendor whose initial design included nozzles has contracted for a commercial system with rotary atomizers (94). Only two of the commercial systems and one of the two full-scale demonstration systems have nozzles but these figures also reflect the preponderance of commercial systems supplied by the Joy/Niro consortium, whose design has a rotary atomizer.

Rotary atomizers consist of a rotor--called a wheel--that is mounted on a vertical shaft driven by a motor-gearbox system. In FGD applications, typical wheels are 8 to 16 inches in diameter and rotate at 10,000 to 20,000 rpm, producing tip speeds of 35,000 to 50,000 ft/sec (93). Absorbent liquid is introduced through an opening in the top of the wheel around the shaft and flows through interior cavities to orifices in the periphery, where it is sheared into droplets as it is expelled. In conventional spray dryers, one atomizer is mounted on the vertical axis at the top of a cylindrical vessel and the gas flows downward around it through the spray. This arrangement places severe demands on the rotary atomizers because of the large flue gas volumes and absorbent flow rates of typical full-scale absorber trains. Very large atomizers, with flow rates well over 100 gal/min in some cases, are required and the flue gas flow has to be modified to achieve effective dispersion of the droplets in the flue gas. The aerodynamic properties of the droplets make them lose velocity very rapidly, forming a cloche-like spray pattern that has at most a diameter of 20 to 30 feet, making it impossible to span the width of full-sized absorbers with a single atomizer and conventional downward gas flow. Some vendors who use rotary atomizers have resorted to installing three or four in each absorber. Others use a single atomizer and introduce a portion of the flue gas below the atomizer, directed upward, to achieve satisfactory mixing. The RI design, for example, has three 300-hp atomizers with wheels less than a foot in diameter and the flue gas is ducted downward around each atomizer (87). The Joy/Niro design, typified by the demonstration system at the Northern States Power Company's Riverside Station (88), has a single 700-hp atomizer with a wheel over a foot in diameter and the flue gas is directed both downward around the atomizer and upward toward the spray. Both have operated successfully and the relative merits of the designs, if any, are not resolved.

The 2-fluid nozzle atomizers have much smaller capacities and consume more power than rotary atomizers, but they are arguably simpler mechanically and more flexible in design and use. The absorbent liquid is combined with air in a chamber in the nozzle and ejected from the nozzle at a high velocity, forming a long narrow plume of turbulent gas and suspended droplets. They commonly have a liquid capacity of about 5 gal/min and operate at gas pressures up to about 100 lb/in^2 (93). Because of the small capacity, at least several must be used in each absorber so the absorber configuration is less important. The B&W system at the Laramie River Station has 15 nozzles in each absorber (83) and the Flakt, Inc., demonstration system at the Pacific Power and Light Company's Jim Bridger Station (83), which has a cylindrical absorber, has 10 nozzles.

Except for the soda ash system at the Coyote Station (which is a direct descendent of RI's sulfur-producing process, based on soda ash), lime has been used exclusively as the absorbent for commercial utility systems. Lime is widely available and with some exceptions in specific areas, it is the least expensive material that is sufficiently reactive for spray dryer FGD. In addition, it produces the now-familiar and relatively insoluble calcium salt waste in dry form that is produced as a slurry in wet-limestone and lime FGD processes. Other absorbents such as magnesium oxide have been evaluated in pilot plants (93).

Unlike many materials that are dried in a spray dryer--in which the drying mechanism tends to produce a porous expanded or hollow particle--the dried particles of lime tend to be dense and have hard surfaces that inhibit reaction of SO_2 with lime in the core (15). This places emphasis on the fineness of the lime particles and has led to more extensive feed preparation techniques than are common in wet FGD processes. Ball mill slakers are used in eight of the commercial utility systems, apparently because tests by some vendors have convinced them of their value. Ordinary paste and detention slakers are also used, however. In addition to producing a very fine particle size, it is also necessary to remove grit and lumps that tend to plug the atomizers.

The presence of fly ash in the absorbent slurry has been found to increase the reactivity of the lime, presumably by providing a large surface area upon which the lime particles can deposit and remain available to SO_2 (95). The effect occurs with acidic fly ash typical of eastern coals and is independent of the effects of increased alkalinity from the fly ash (92). For this reason, and to take advantage of unreacted lime and the alkalinity of the fly ash itself, most vendors recycle a portion of the waste to the spray dryer (83).

The absorbent slurry composed of fresh lime and reslurried waste is normally maintained at a constant solids content and the addition rate is varied to provide the desired SO_2 removal efficiency. Normally, L/G ratios of 0.2 to 0.3 gal/kft^3 are used. Additional water is added to the atomizer to control the outlet flue gas temperature. The maximum slurry solids with fly ash is apparently about 35%, somewhat lower for slurries composed wholly of fresh lime, above which the slurry becomes too viscous and abrasive to pump.

Stoichiometries are generally expressed in mols of absorbent per mol of inlet SO_2 rather than mols of SO_2 removed, as is common in wet FGD processes. Furthermore, fly ash alkalinity is regarded as an important factor in subbituminous coal and lignite applications; SO_2 removals of 65% have been reported using alkaline fly ash alone (96). Stoichiometry alone is thus not a reliable indication of SO_2 removal efficiency or absorbent utilization (that is, the percentage that actually reacts with the SO_2). Reported design stoichiometries (83) for lime-based commercial systems range from 0.88 to 1.5 mol $Ca(OH)_2$/mol SO_2 inlet, with little relationship to design SO_2 removal. The 0.88 stoichiometry corresponds to an SO_2 removal of 85%, for example, while the 1.5 stoichiometry corresponds to an SO_2 removal of 86%. In general, the stoichiometries are higher for higher-sulfur coal applications.

The most important operating condition is the temperature change in the flue gas caused by evaporation of the absorbent water. Normally, flue gas from the boiler air heater has a temperature of about 300°F and a water saturation temperature of roughly 125°F. The spray dryer must operate within this range to avoid producing moist particulates which can deposit in the spray dryer and ducts (a "wet bottom" condition) or clog the baghouse. The approach to saturation--which for a constant absorbent rate is determined by the solids content of the absorbent--is important because SO_2 removal efficiency increases rapidly as the temperature of the flue gas leaving the spray dryer approaches its saturation temperature, allowing a longer effective liquid phase to exist in the absorbent droplets.

Commercial system design thus far allows for a 15°F to 40°F approach to saturation, with about 20°F most common. A close approach to saturation increases absorbent utilization while increasing control problems and the possibility of upsets. It is most important for high-sulfur coals requiring high SO_2 removal efficiencies--and for which absorbent costs are a critical factor. A close approach in these cases may also be necessary to introduce sufficient absorbent since the solids content of the absorbent liquid is limited by absorbent solubility or slurry viscosity.

WET-LIMESTONE FLUE GAS DESULFURIZATION

Since it was first used by U.S. utilities over 15 years ago, wet-limestone scrubbing has become the standard method of utility FGD in the United States. In 1981, 39 systems, 42% of all utility FGD systems in operation, were limestone systems and among those being constructed or in the planning stage, 46% were limestone systems (97). After a long and sometimes difficult evolution, limestone FGD has become a mature and reliable technology (98). This evolution has produced a spectrum of processes differing appreciably in design and operating philosophies. All, however, share the restrictions placed on them by the use of this cheapest of available absorbents: large liquid circulation rates of an abrasive slurry, restriction to a narrow range of operating conditions to avoid scaling and plugging, and the production of an intractable waste.

The process used in this study represents two current trends in limestone FGD: the use of a simplified absorber design and the use of forced oxidation to improve the waste-handling properties. The process is one of several variants that were extensively evaluated during tests sponsored by EPA at the Shawnee test facility operated at TVA's Shawnee Steam Plant (99).

Most limestone FGD systems have absorbent liquids that contain between 5% and 15% solids that consist of unreacted limestone, calcium sulfite, calcium sulfate, and sometimes fly ash. The relatively insoluble limestone (the solubility constant is about 10^{-9}, several orders of magnitude lower than other FGD absorbents) makes the use of a stoichiometric excess of limestone necessary. Normally, the stoichiometric ratios lie in the range of 1.1 to 1.6 mol $CaCO_3$/mol SO_2 removed. The operating pHs usually range from 5 to 6 at the absorber inlet to 4.5 to 5.5 at the absorber outlet, depending in large part on the stoichiometry. Because of the slow dissolution rate of the limestone, limestone processes require a large hold tank with a hold time of at least several minutes to allow time for a portion of the dissolution and precipitation reactions to occur outside the absorber itself (100).

Regardless of the operating conditions, SO_2 gas-liquid mass transfer rates are low compared with most FGD processes and intimate gas-liquid contact is necessary. This requires either multistage absorbers, absorbers with complex internals, or very large L/G ratios. All of these are widely used, typified by venturi-spray tower combinations, mobile-bed absorbers, and spray towers with high L/G ratios (98). In recent years, with the maturation of limestone FGD technology, there has been a trend toward simple absorbers such as spray towers (101), motivated apparently by the lower capital costs and maintenance requirements. There is relatively little difference in operating costs among the different approaches because fan costs for high-pressure drop designs tend to be offset by pumping costs for high L/G designs (14).

Chemistry

The chemistry of limestone FGD--which is surprisingly complex, with many interrelated factors that differ in importance depending on the operating conditions--has been discussed at length (102). The primary alkaline species that react with dissolved SO_2 to form SO_3^{-2} and HSO_3^- are CO_3^{-2} and HCO_3^- from limestone dissolution and SO_3^{-2} from calcium sulfite dissolution. These, along with several minor species, are referred to as the total dissolved alkalinity (103). The level of dissolved alkalinity can be increased somewhat by the addition of cations that form more soluble salts than calcium, such as magnesium (103), or by the use of a buffer, such as adipic acid (104). The forms of all the sulfite and carbonate species are pH dependent and the pH of the absorbent is vitally important because of its effects on SO_2 removal efficiency and the precipitation of solids. The relative importance of the many reactions involved is thus dependent on the operating conditions--limestone stoichiometry, SO_2 absorption rate, L/G ratio, hold times, and others.

Within limits, a high pH is desirable because of the higher SO_2 absorption rate. At pHs above about 6, however, the formation of carbonate scale may cause operating problems (99). Some decrease in pH, as the absorbent liquid passes through the absorber, is also desirable because it causes the absorbed SO_2 to exist as the soluble HSO_3^- rather than to precipitate as calcium sulfite, which can form muddy accumulations that plug the absorber. Too large a decrease, however, may cause an inert coating to form on the limestone particles that prevents their dissolution. A low pH leaving the absorber may also be desirable if forced oxidation is incorporated in the process. A low pH may also affect the formation of sulfate scale: low pHs favor the dissolution of calcium sulfite rather than limestone as the source of dissolved alkalinity. This produces twice as many mols of calcium ions per mol of SO_2 absorbed as limestone dissolution, which under some conditions, can produce high levels of gypsum $CaSO_4 \cdot 2H_2O$ supersaturation. There are, however, other methods of dealing with sulfate scale.

The absorbent liquid leaving the absorber consists of a disequilibrium mixture of dissolved sulfite and sulfate species, supersaturated in respect to calcium sulfite and gypsum, and solid limestone, calcium sulfite, and gypsum. The limestone continues to dissociate and the calcium salts precipitate as the pH slowly rises. Addition of fresh limestone slurry accelerates the precipitation reactions which, after several minutes or more, have progressed sufficiently for the regenerated absorbent to be reused in the absorber.

Because of the intimate gas-liquid contact necessary for SO_2 absorption, oxygen absorption is also high in limestone FGD and the proportion of the SO_2 absorbed that is oxidized to gypsum is correspondingly high compared with most FGD processes. For high-sulfur coals, oxidation rates vary widely, with oxidation of about one-third of the total SO_2 absorbed being representative. For low-sulfur coal applications, oxidation rates in excess of 90% are possible. These increase the problems caused by gypsum scaling that are common to calcium-based systems but which have now largely been controlled.

The common gypsum scaling, which is a problem in many industrial water systems as well as calcium-based FGD systems, is a result of the tendency of gypsum to form supersaturated solutions from which rapid precipitation can occur when a critical concentration is reached. Under some conditions, this precipitation can form rock-like deposits on the surfaces of the equipment. Because of the solubility relationships of calcium sulfite and gypsum, it is impractical to operate limestone FGD systems below gypsum saturation; most, in fact, operate up to about 1.3 times gypsum saturation, a level below which experience has shown gypsum scaling to be minimized. Among the methods of controlling the gypsum saturation level are the provision of abundant seed crystals, providing sufficient reaction time outside the absorber for limestone dissolution and gypsum precipitation, reducing the quantity of SO_2 absorbed per volume of absorbent liquid (the "make per pass"), and operating in a pH and stoichiometry range that does not create large increases in the calcium ion concentration in the absorber.

Forced Oxidation

The ease with which sulfites in solution are oxidized by dissolved oxygen creates scaling problems but it also affords a means of reducing one of the major problems in limestone FGD: handling and disposing of the intractable high-sulfite waste. High-sulfite wastes cannot be dewatered sufficiently to form a stable solid; they must either be impounded or treated with stabilizing material. Chemically precipitated gypsum, in contrast, is a sand-like material that can easily be dewatered to a stable solid that can be disposed of in a landfill without further treatment (105). There is also the prospect, at least technically feasible, of disposing of gypsum as a byproduct to replace natural gypsum in industrial uses (106).

Forced oxidation has been widely adopted in Japan as a means of improving waste properties and of producing a byproduct (107). The complex 2-stage processes used there, however, did not prove attractive to U.S. utilities. In the United States, institutional and industrial studies of forced oxidation in conventional limestone processes were begun in the mid-1970s. The most fully documented are the pilot-plant tests begun by EPA in 1975 and continued at the Shawnee test facility from 1976 through 1979 (99). During the same period, forced-oxidation versions of limestone FGD processes were developed by several vendors of FGD systems (108). In 1983, there were about a dozen full-scale utility systems in operation or under construction (108).

The forced-oxidation method consists of sparging air into the absorbent liquid so that sufficient oxygen is absorbed to oxidize the sulfites in the liquid to sulfate, which reacts with the calcium already present to precipitate gypsum. Normally, this is done in the absorber liquid loop, in one of the absorber hold tanks, or in an additional tank in the loop. Forced oxidation of a bleedstream is also used (109) but this requires the presence of an additive such as magnesium or adipic acid to increase the concentration of dissolved sulfites because the pH increases as oxidation progresses reduce the oxidation rate below practical levels (99). Usually a simple perforated piping system and a low-pressure air supply serves as the sparging system. The tank is agitated to provide sufficient air-liquid content and keep the solids in suspension (other methods using air ejectors and agitation by air injection are also used less frequently). The air rate, expressed in atomic equivalents of oxygen per mol of SO_2 absorbed, can be as low as 1.5 lb atoms O/lb mol SO_2 but 2 to 5 is more common. The essential requirements are a pH of about 5.5 or less to ensure that the sulfite is in solution as HSO_3^- rather than in solid form as calcium sulfite and that there is adequate air-liquid contact time.

Forced oxidation does not decrease the SO_2 absorption efficiency because of the depletion of dissolved alkalinity provided by SO_3^{2-}, probably because the absence of sulfite species improves the gas-liquid mass transfer rate (110). It has been known for some time (111) that forced oxidation decreases the gypsum supersaturation level and the scaling potential, rather than increasing it as might be supposed. The precipitation of gypsum at moderate supersaturation levels is enhanced because of the abundant seed crystals in the liquid. The high degree of agitation normally used may also enhance nucleation of new crystals (112).

PREMISES

The premises used in this study were developed by TVA to evaluate the economics of coal-fired power plant emission control processes. The design premises quantify SO_x, NO_x, and particulate emissions from a typical modern coal-fired power unit and establish representative design and operating conditions required to determine emission control economics. The economic premises define the procedures for determining the capital investment and annual revenue requirements based on regulated utility economics. The premises are based on projects with a 1981 to 1983 construction period and a 1984 startup. Capital investment is based on mid-1982 costs and annual revenue requirements are based on mid-1984 costs.

DESIGN PREMISES

Coal Premises

The coals used are an eastern bituminous coal containing 3.5% sulfur and a western subbituminous coal containing 0.7% sulfur, both on a dry basis. The properties of these coals are based on composites of samples representing major coal production areas (113,114,115). The eastern bituminous coal has a heating value of 11,700 Btu/lb and an ash content of 15.1% as fired. The heating value and ash content of the western subbituminous coal are 8,200 Btu/lb and 6.3% as fired. The compositions of the coals are shown in Table 2.

Ash compositions are considered to be typical of the coals used. The compositions are not qualified in terms of physical and chemical behavior, with the exception of calcium content. Both ashes are assumed identical in handling properties until wetted. The eastern coal ash is assumed to have no cemetitious properties affecting handling and disposal site emplacement. The western coal ash is assumed to have self-hardening characteristics that affect handling and emplacement within a few hours after being wetted (16).

Power Plant

The power plant site is in the north-central region (Illinois, Indiana, Ohio, Michigan, and Wisconsin). The location represents an area in which coal-fired power plants burning coals of diverse type and source are situated (12,116). The design is based on standard design practices (19,20) and current trends in utility boiler construction (18). The base case power unit is a new, single 500-MW, balanced-draft, pulverized-coal-fired, dry-bottom boiler. Heat rates are based on power unit size as shown in Table 3. To provide equitable comparisons, the power units are not derated for energy

TABLE 2. COAL COMPOSITIONS

Coal	Sulfur, %	Ash, %	Moisture, %	Heat content, Btu/lb	Ultimate analysis C,%	H,%	O,%	N,%	Cl,%
(As-Fired Basis)									
Eastern bituminous	3.36	15.14	4.0	11,700	66.7	3.8	5.6	1.3	0.1
Western subbituminous	0.48	6.30	29.3	8,200	49.0	3.5	10.7	0.7	0.02
(Moisture-Free Basis)									
Eastern bituminous	3.50	15.7			69.5	4.0	5.8	1.4	0.1
Western subbituminous	0.68	8.9			69.3	5.0	15.1	1.0	0.02

consumption by the emission control systems evaluated. Instead, the energy requirements are charged as independently purchased commodities. Cost estimates are based on a single power unit independent of other units at the site. Power unit size case variations consist of similar 200- and 1,000-MW units. The emission control systems are assumed to be installed during construction of the power plant and are assumed to have a 30-year life and to operate at full load for 5,500 hours a year. This operating schedule is equivalent to a total lifetime operation of 165,000 hours.

TABLE 3. POWER UNIT OPERATING TIME AND HEAT RATE

Power unit size, MW:	200	500	1,000
Remaining life, yr	30	30	30
Full load, hr/yr	5,500	5,500	5,500
Heat rate, Btu/kWh			
Bituminous coal	9,700	9,500	9,200
Subbituminous coal	10,700	10,500	10,200

Flue Gas Compositions

Flue gas compositions are based on combustion of pulverized coal, assuming a total air rate equivalent to 139% of the stoichiometric requirement (the air required for combustion of carbon, hydrogen, and sulfur). This includes 20% excess air to the boiler and 19% additional air leakage to the flue gas in the air heater. It is assumed that 80% of the ash present in each coal is emitted as fly ash and the remaining 20% as bottom ash, with no adjustments for pulverizer rejects or slagging and fouling losses. For the bituminous coal, 92% of the sulfur is emitted as SO_x, while 85% of the sulfur is emitted as SO_x for the western subbituminous coal. The remaining sulfur is removed in the bottom ash and fly ash. No loss of sulfur in the pulverizers is assumed. Three percent of the sulfur emitted as SO_x is SO_3 and the remainder is SO_2. The base case flue gas compositions and flow rates based on combustion of each of the coals assumed for the study are shown in Tables 4 and 5.

Environmental Standards

The NSPS established by EPA in 1979 governing emissions from new coal-fired utility plants specify a maximum emission for particulate matter, SO_2, and NO_x emissions based on heat input. These emission standards are shown in Table 6.

TABLE 4. FLUE GAS COMPOSITION
FOR 3.5% SULFUR EASTERN BITUMINOUS COAL

Component	Vol. %	Lb-mol/hr	Lb/hr
N_2	74.85	118,700	3,326,000
O_2	3.27	5,178	165,700
CO_2	14.22	22,550	992,300
SO_2	0.24	380	24,320
SO_3	0.01	12	940
NO	0.04	59	1,766
NO_2	0.00	3	142
HCl	0.01	12	418
H_2O	7.36	11,660	210,200
	100.00	158,600	4,722,000
Fly ash			50,050
Total			4,772,000

Sft3/min (60°F) = 1,003,000
Aft3/min (705°F) = 2,247,000

Fly ash loading

	Gr/sft3
Wet	5.82
Dry	6.28

Basis: 500-MW boiler, flue gas conditions after economizer at 705°F.

In this study, the particulate matter and SO_2 removal efficiencies for all cases are designed to meet these NSPS. The SO_2 and particulate matter removal efficiencies are tabulated in Tables 7 and 8. The boilers being constructed now are capable of meeting the current NSPS for NO_x emissions without flue gas treatment. The SCR processes are capable of achieving 80% to 90% NO_x reduction efficiencies. Therefore, in this study, the NO_x emission level from the boiler for most cases is assumed to be equal to the existing NSPS limitations and the NO_x reduction efficiency with SCR is an additional 80%. This produces a stack NO_x emission of 0.12 lb/MBtu for the bituminous coal and 0.10 lb/MBtu for the subbituminous coal. A case variation of 90% NO_x reduction efficiency is examined to illustrate the effects of varying NO_x reduction efficiency on the economics of the systems evaluated.

TABLE 5. FLUE GAS COMPOSITION

FOR 0.7% SULFUR WESTERN SUBBITUMINOUS COAL

Component	Vol. %	Lb-mol/hr	Lb/hr
N_2	70.22	134,900	3,780,000
O_2	3.07	5,894	188,600
CO_2	13.59	26,120	1,150,000
SO_2	0.04	79	5,063
SO_3	0.00	2	196
NO	0.03	54	1,627
NO_2	0.00	3	131
HCl	0.00	4	132
H_2O	13.05	25,070	451,700
	100.00	192,100	5,577,000
Fly ash			32,640
Total			5,610,000

Sft3/min (60°F) = 1,216,000
Aft3/min (780°F) = 2,900,000

Fly ash loading

	Gr/sft3
Wet	3.13
Dry	3.60

Basis: 500-MW boiler, flue gas conditions after economizer at 780°F.

Waste disposal sites are assumed to be governed by nonhazardous solid waste regulations. Landfill disposal is used for the ash, FGD waste, and NO_x catalyst.

TABLE 6. 1979 REVISED NSPS EMISSION STANDARDS

SO_2

70% SO_2 removal (minimum) to a maximum SO_2 emission of 0.6 lb SO_2/MBtu
0.6 lb SO_2/MBtu maximum emission up to 90% SO_2 removal
90% SO_2 removal (minimum) to a maximum SO_2 emission of 1.2 lb SO_2/MBtu
1.2 lb SO_2/MBtu maximum emission

NO_x

Bituminous coal - 0.6 equivalent lb NO_2/MBtu
Subbituminous coal - 0.5 equivalent lb NO_2/MBtu
Lignite - 0.6 equivalent lb NO_2/MBtu

Particulate

0.03 MBtu

TABLE 7. SO_2 EMISSION CONTROL REQUIREMENTS

Coal	Case No.	Equivalent SO_2 content of coal, lb SO_2/MBtu	Overall equivalent SO_2 removal efficiency, %	Equivalent SO_2 removal required in FGD system, %[a]	Controlled outlet emission, lb SO_2/MBtu
Eastern bit., 3.5% S	1	5.74	89.6	88.7	0.60
Western subbit., 0.7% S	2,3	1.17	70.0	64.7	0.35

[a] Based on FGD as the only SO_2 control device and the previously defined sulfur retention in the ash.

TABLE 8. PARTICULATE MATTER EMISSION CONTROL REQUIREMENTS

Coal	Case No.	Fly ash[a] content of flue gas entering ESP or baghouse, lb/MBtu	Fly ash[a] removal required in ESP or baghouse process, %	Fly ash controlled outlet emission, lb/MBtu
Eastern bit., 3.5% S	1	10.54	99.7	0.03
Western subbit., 0.7% S	2	21.43[a]	99.9	0.03
Western subbit., 0.7% S	3	6.87	99.6	0.03

a. Particulate matter in case 2 flue gas is a combination of fly ash and lime spray dryer solids.

NO_x Control Process

The SCR NO_x control process is a generic design based on information from several vendors. Two reactor trains are used for the 200-MW and 500-MW boilers and four trains are used for the 1,000-MW boiler. Catalyst replacement is assumed to take place during scheduled boiler outages and no spare reactor trains are provided. Spare ammonia vaporization equipment is provided. The catalyst is assumed to have a 1-year life. The spent catalyst is assumed to be a nonhazardous waste that can be disposed of in the facility landfill.

The costs for an economizer bypass are also included in the costs. This is used at low boiler loads (less than 60% of capacity) to maintain a sufficiently high flue gas temperature in the reactors. Air heater modifications to minimize the effects of ammonia salts on the air heaters are also included. These consist of an increase in the size of the air heaters, use of a thicker gauge corrosion-resistant alloy for the elements, and provisions for additional sootblowers and water washing. The additional costs--compared with the costs of conventional air heaters--are included in the NO_x process costs. A bypass duct around the SCR reactor is included for maximum system flexibility.

Separate booster fans are not used. Instead, the boiler ID fans are increased in size to compensate for the pressure drop in the NO_x control system and the incremental costs of this increase are included in the NO_x

process costs. (It is possible in some cases that boiler modifications would be necessary to reduce the possibility of implosions caused by the higher overall pressure drops. These are not included because they are highly variable and site specific.)

Several other possible effects of the NO_x process that have economic implications are not included because they are undefined. These include the effects of ammonia and ammonia salts on downstream equipment other than the air heater and the possible need of additional waste water processing and waste disposal requirements because of the presence of soluble nitrogen compounds.

FGD Process

The limestone and spray dryer FGD processes are generic designs. The limestone process is based on data developed at the Shawnee test facility (99,117), industry information, and vendor information. The spray dryer process is based on vendor information and published data. Both represent current trends in industry practice.

The FGD systems consist of multiple absorber trains supplied by a common plenum into which the boiler ID fans (or cold-side ESPs) discharge. Spare trains are provided in all cases to permit the use of an emergency bypass, as specified in the 1979 NSPS (32). The number of trains for each of the cases and power unit sizes evaluated is shown in Table 9. The emergency bypass consists of two ducts, each sized for 25% of the flue gas scrubbed, that connect the ends of the inlet plenum with the stack plenum. If partial scrubbing is used and some flue gas is normally bypassed, the size of these ducts is increased proportionally and they are also used for the normal bypass.

TABLE 9. NUMBER OF FGD TRAINS

	Case 1		Case 2		Case 3	
Unit size, MW	Operating	Spare	Operating	Spare	Operating	Spare
200	2	1	2	1	2	1
500	4	1	3	1	3	1
1,000	8	2	6	2	6	2

For the limestone FGD processes, each train is equipped with an ID booster fan to compensate for the pressure drop in the FGD system. For the spray dryer FGD process, ID booster fans downstream from the baghouses serve to compensate for the pressure drop in both the spray dryer system and the baghouse.

For low-sulfur coal applications, the absorbers are designed for the highest practical removal efficiency (90% for limestone FGD and 73% for spray dryer FGD) and a portion of the flue gas is bypassed to reduce reheating requirements. This is more economical than processing all of the flue gas at lower SO_2 removal efficiencies. For the 0.7% sulfur coal using limestone FGD, 28% of the flue gas is bypassed; for the same coal using spray dryer FGD, 12% is bypassed.

The limestone FGD process has spray tower absorbers constructed of rubber-lined carbon steel. They are equipped with a presaturator section at the inlet in which the flue gas is sprayed with absorbent liquid to cool it to 127°F. Each absorber has a horizontal mist eliminator with fiberglass chevron vanes that reduces the entrained moisture to 0.1% by weight of the flue gas.

For the limestone FGD processes, indirect steam reheat of the scrubbed gas is provided as necessary to provide a flue gas temperature of 175°F in the stack. The size of the reheater is based on a scrubbed gas temperature of 125°F and a bypassed gas temperature of 300°F. The heat of compression in the ID booster fans is also included in the determination of reheater size. The reheater tubes are Inconel 625 at flue gas temperatures below 150°F and Cor-Ten steel above 150°F. It is designed for a flue gas velocity of 25 ft/sec.

The spray dryer process has cylindrical absorbers with conical bottoms that are equipped with single rotary atomizers. They are constructed of unlined carbon steel. The spray dryers are enclosed in a prefabricated metal building to reduce seasonal temperature variations.

Square or rectangular insulated ductwork is used. At temperatures below 150°F, it is constructed of stainless steel. At temperatures above 150°F, it is constructed of Cor-Ten steel. All ducts are designed for a flue gas velocity of 50 ft/sec.

Particulate Control Process

The particulate control process consists of the ESPs or baghouses, all hoppers associated with ash collection on the boiler (bottom ash, economizer ash, and air heater ash) and emission control equipment (NO_x process reactors and the ESPs or baghouses), a pneumatic conveying system and storage system for the dry particulates, a hydraulic conveying system, and dewatering system for the bottom ash. All designs are based on standard industry practice and commercially available equipment.

The ESPs and baghouses are standard designs based on current industry practice and vendor information. Two ESPs in parallel are used for the 200-MW and 500-MW boiler sizes and four trains in parallel are used for the 1,000-MW boiler. All dry particulate hoppers are a double-vee bottom design with 55-degree slopes and are insulated and electrically heated. Hoppers are provided on all boiler and emission control equipment upstream of the ESPs or baghouses. The bottom ash system is a standard utility design with a double-vee, water-filled hopper; ejector and centrifugal pumps; and a bin-type dewatering system. Vacuum fly ash conveying systems are used for systems with ESPs. Vacuum-pressure systems are used with systems with baghouses because of the larger volumes and larger number of hoppers involved.

All hoppers are designed for a 12-hour capacity to allow intermittent removal of solids. Storage facilities (bottom ash bins and fly ash silos) are designed for a 3-day capacity to permit waste transportation operations on a 5-day week.

Solids Disposal

The solids generated by the emission control processes (FGD solids, fly ash, bottom ash, and spent catalyst) are disposed of in a common landfill one mile from the facility. Sufficient land is provided for disposal for the 30-year life of the facility. The disposal site is assumed to be an area of low relief with sufficient soil for landfill preparation and reclamation.

Square area-type landfills with a 20-foot-high perimeter and a 6-degree cap are used, as shown in Figure 2. After topsoil removal, the landfill area is lined with 12 inches of clay (assumed available onsite). A French drain system (perforated pipe surrounded by gravel) and 24 inches of bottom ash are added on top of the clay. The bottom ash layer and French drain system allow the seepage to be collected separately from runoff and treated for pH adjustment.

Land requirements include the landfill, catchment basin, equipment storage area, topsoil storage area, and a 50-foot perimeter of undisturbed land. Costs for access roads, a 6-foot security fence around the total landfill area, security lighting, topsoil stripping and replacement, and revegetation are included. One upstream and three downstream groundwater monitoring wells are also included.

All mobile equipment involved in loading and transporting the solid wastes from the in-process storage area, as well as working the landfill, is included in the landfill equipment. Mobile equipment and landfill requirements are based on the quantity, moisture content, and bulk density of the solid wastes. The dry bulk densities and water contents for each solid waste are shown in Table 10.

Raw Materials

Raw materials capacity is normally 30 days unless process or industry practice differs. Standard raw material characteristics are shown in Table 11.

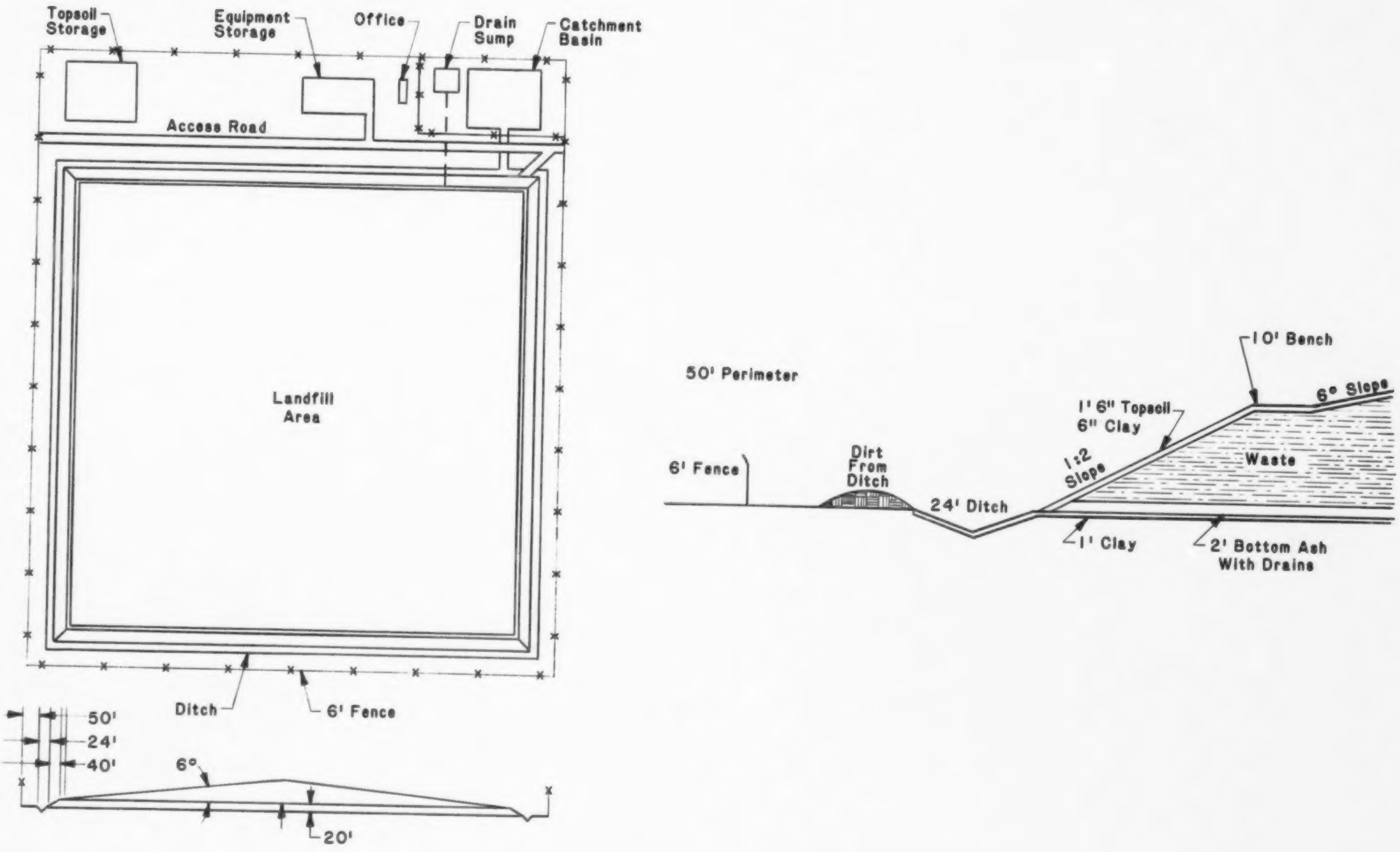
Topsoil Storage
Equipment Storage
Office
Drain Sump
Catchment Basin
Access Road
Landfill Area
Ditch
6' Fence
50'
24'
40'
6°
20'
50' Perimeter
6' Fence
Dirt From Ditch
24' Ditch
1:2 Slope
1' 6" Topsoil
6" Clay
10' Bench
6° Slope
Waste
1' Clay
2' Bottom Ash With Drains

Figure 2. Landfill plan and construction details.

TABLE 10. SOLID WASTE PERCENT MOISTURE AND DENSITY

Type of solid waste	% moisture			Dry bulk density, lb/ft3					
				Truck transport			Landfill		
	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3	Case 1	Case 2	Case 3
Bottom ash	10	10	10	45	51	51	75	85	85
Fly ash	10	-	0	55	-	62	85	-	95
FGD	15	-	15	55	-	55	85	-	85
Fly ash/FGD	-	0	-	-	62	-	-	95	-
NO_x catalyst	0	0	0	48	48	48	48	48	48

TABLE 11. RAW MATERIAL CHARACTERISTICS

	Size as received	Ground size	Analysis	Bulk density, lb/ft
Limestone	0 x 1-1/2 in.	90% to pass 325 mesh	95% $CaCO_3$[a] 0.15% MgO 4.85% inerts 5 lb H_2O/100 lb dry limestone	95
	Fineness of grind index factor = 5.7 Hardness of work index factor = 10			
Lime (pebble)	3/4 x 1-1/4 in.	-	95% CaO[a] 0.15% MgO 4.85% inerts 5 lb H_2O/100 lb dry lime	55
Sulfuric acid		-	98% H_2SO_4	-
Caustic soda		-	50% NaOH	-
Ammonia		-	$\geq$99.5% NH^3 82.2% N	-

a. Limestone and lime analysis on a dry basis. H_2O is based on pounds of dry limestone or lime.

ECONOMIC PREMISES

Schedule and Cost Factors

A 3-year construction period, from early 1981 to late 1983, is used. Mid-1982 costs are used for capital investment and mid-1984 costs are used for annual revenue requirements. These costs represent the midpoint of construction expenditures in 1982 and the midpoint of the first year of operation in 1984. Costs are projected from Chemical Engineering magazine cost indexes, as shown in Table 12, using standard estimating techniques (118,119). Frequently used costs are shown in Table 13. In some instances, cost-scaling factors based on gas and product rates are used to calculate values at conditions other than the base case.

TABLE 13. COST FACTORS

Project Timing	
Start	January 1981
End	December 1983
Midpoint	Mid-1982
First-year operation	1984
1984 Utility Costs	
Electricity	\$0.037/kWh
Steam	\$2.50/klb, \$3.30/MBtu
Diesel fuel[a]	\$1.60/gal
Filtered river water	\$0.14/kgal
1984 Labor Costs	
Process operating labor	\$15.00/man-hr
Waste disposal labor	\$21.00/man-hr
Analyses labor	\$21.00/man-hr
1984 Raw Material Costs	
Limestone	\$8.50/ton (95% $CaCO_3$, dry basis)
Lime	\$75/ton (pebble, 95% CaO, dry basis)
Catalyst	\$23,558/ton (\$568/ft3)
Ammonia	\$155/ton
Sodium hydroxide	\$300/ton
Sulfuric acid	\$65/ton
1982 Land Cost	\$5,000/acre

These cost factors are based on a north-central plant location.

---

[a] Cost is based on wholesale price of barge-load quantities. Road taxes are not included.

TABLE 12. COST INDEXES AND PROJECTIONS

Year:	1978	1979	1980[a]	1981[a]	1982[a]	1983[a]	1984[a]
Plant	218.8	238.7	257.8	277.1	297.9	320.2	342.6
Material[b]	240.6	264.4	288.2	311.2	336.1	363.0	388.4
Labor[c]	185.9	194.9	210.5	227.3	245.5	265.2	283.7

a. TVA projections.

b. Same as "equipment, machinery, supports" Chemical Engineering index.

c. Same as "construction labor" Chemical Engineering index.

Capital Cost Estimates

The capital investment estimates are divided into three major sections: direct investment, indirect investment, and other capital investment. The sum of direct and indirect investments is the fixed investment.

Direct Investment--

Direct investment consists of the installed costs of all process equipment, including provisions for services, utilities, and miscellaneous; and waste disposal investment. Installation costs are estimated by major processing area and include charges for all piping, foundations, excavation, structural steel, electrical equipment, instruments, ductwork (all flue gas ductwork is included in the gas-handling area), paint, buildings, taxes, freight, and a premium for 7% overtime construction labor.

Service facilities such as maintenance shops, stores, communications, security, offices, and road and railroad facilities are estimated or allocated on the basis of process requirements. Included in the utilities costs are necessary electrical substations and facilities for process water, fire and service water, instrument air, chilled water, inert gas, and compressed air. Services, utilities, and miscellaneous are 6% of the total process capital.

All equipment and direct construction costs associated with the landfill are included in waste disposal costs. All mobile equipment involved in loading and transporting the waste from the in-process storage area, as well as working the landfill, is included in landfill equipment. The solid wastes from all control processes are disposed of in a common landfill. The landfill costs are prorated to each of the individual processes based on the volume of solid waste from each process. The sum of total process capital, services, utilities, and the waste disposal cost is the total direct investment.

Indirect Investment--

Indirect capital costs cover fees for engineering design and supervision, architect and engineering contractor, construction expense, contractor fees, and contingency. Listed in Tables 14 and 15 are the percentages used to calculate these costs.

TABLE 14. INDIRECT CAPITAL COST FACTORS

	Indirect investment					
	Process[a]			Landfill[b]		
	1,000 MW	500 MW	200 MW	1,000 MW	500 MW	200 MW
Engineering design and supervision	6	7	8	2	2	2
Architectural and engineering	1	2	3	1	1	1
Construction expense	14	16	18	7	8	9
Contractor fees	4	5	6	4	5	6
	25	30	35	14	16	18

a. Percentage of process direct investments, excluding landfill.
b. Percentage of landfill equipment and construction.

TABLE 15. CONTINGENCY AND ALLOWANCE FOR STARTUP AND MODIFICATION COST FACTORS

Process type[a]	Contingency, % of direct and indirect investment	Allowance for startup and modification, % of fixed investment
NO_x-SCR	20	10
SO_x-limestone slurry	10	8
SO_x-lime spray dryer	20	10
Particulate-ESP/baghouse	20	10
Waste disposal[b]		
Landfill	20	0

a. Percentage of only process costs.
b. Percentage of only landfill costs.

Other Capital Investment--

The allowance for startup and modifications is applied as a percentage of the total fixed investment. Since the startup and modification costs for the waste disposal area are assumed to be negligible, this allowance is calculated using only the process fixed investment. The percentages used to calculate these costs are listed in Tables 14 and 15.

The cost of borrowed funds (interest) during construction is 15.6% of the fixed investment (both process and waste disposal). This factor is based on an assumed 3-year construction schedule and is calculated with a 10% weighted cost of capital, with 25% of the construction expenditures in the first year, 50% in the second year, and 25% in the third year of the project. Expenditures in a given year are assumed uniform over that year. Startup costs are assumed to occur late in the project schedule so that there are no charges for the use of money to pay startup costs. No royalty charges are included for the limestone or lime FGD processes and for particulate control by either ESP or baghouse. Royalty charges are included for the SCR processes and are based on vendor information.

Land--

All land associated with the process and waste disposal area is charged to the process. The cost of land is 5,000 $/acre.

Working Capital--

Working capital is the total amount of money invested in raw materials, supplies, finished products, accounts receivable, and money on deposit for payment of operating expenses. Working capital is the equivalent cost of 1 month's raw material cost, 1.5 months' conversion cost, 1.5 months' plant and administrative overhead costs (all of the above are shown on the annual revenue requirements sheet), and 3% of the total direct capital investment. One month is defined as 1/12 of annual costs. For the SCR processes, catalyst replacement cost is excluded in calculating the working capital.

Annual Revenue Requirements

Annual revenue requirements account for recovery of various direct and indirect operating and maintenance costs and capital charges. Annual revenue requirements normally vary from year to year as operating and maintenance costs change and capital charges decline. Thus, no single year is necessarily representative of the lifetime costs, and single-year undistorted comparisons cannot be made among processes with different ratios of operating costs to capital charges. In addition, it is necessary to take into account the effect of time on the value of money (i.e., for inflation, the future earning power of money spent, and other factors).

Frequently these factors are accounted for by levelizing (120). Levelization converts all the varying annual revenue requirements to a constant annual value, such that the sum of the present worths of the levelized annual revenue requirements equals the sum of the present worths of the actual annual revenue requirements. The levelized value is calculated by multiplying the revenue requirements for each year by the appropriate present worth factor and summing the present worth values. Then the single present worth value is converted to equal annual values by multiplying the result by the capital recovery factor.

In this study, the operating and maintenance costs are levelized by multiplying the first-year operating and maintenance cost by a levelizing factor. The levelized capital charges are determined by levelizing the percentage of capital investment applied yearly as capital charges. The levelizing factor includes a discount factor reflecting the time value of money and an inflation factor reflecting the effects of inflation during the operating life of the system. The annual discount rate used is 10% and the annual inflation rate used is 6%.

Operating and Maintenance Costs--

Operating and maintenance costs consist of direct costs for raw materials and conversion and indirect costs for overheads. Conversion costs consist of operating labor and supervision, utilities, maintenance, and analysis costs.

Raw materials include consumables required for their chemical or physical properties, other than fuel for the production of heat. In this evaluation, the raw materials consist of limestone, lime, ammonia, catalyst, sulfuric acid, and sodium hydroxide. Raw material costs are determined from vendor quotations or published sources. All costs are delivered costs.

Operating labor and supervision consists of all labor requirements for operation of the equipment and waste disposal facility. The allocation of operating labor and supervision man-hours for the NO_x-SO_2-particulate control system depends on the process complexity, number of process areas, labor intensity of the process, and operating experience. Waste disposal operating labor and supervision depends on the number of equipment operators required to operate the landfill.

Services such as steam, electricity, process water, and diesel fuel are charged under the utilities heading. Costs for steam and electricity are based on the assumption that the required energy is purchased from another source. This simplifying assumption eliminates the need to derate the utility plant. Process water requirements are defined as any water used by the process being evaluated and are determined from the material balance. Steam requirements are for stack gas reheat and sootblowing. Electrical power requirements are determined from the installed horsepower of operating electrical equipment (excluding the horsepower of spared equipment). Each motor in operation is assumed to be operating at rated capacity although this results in higher power consumptions than would actually occur. Electrical requirements are obtained from the equipment list where the motor horsepower is identified, plus an additional amount for functions such as lighting. Diesel fuel is calculated based on the equipment required for the landfill.

Process maintenance costs are calculated as a percentage of direct process investment which varies with the process complexity, process equipment, materials handled, and the power unit size. Waste disposal maintenance costs are 3% of the waste disposal direct investment. The maintenance percentages used for specific processes are shown below. Analysis costs are based on process complexity and are listed as a single entry.

	Maintenance, % of direct investment		
Process type (process only)	1,000 MW	500 MW	200 MW
NO_x-SCR	3	4	5
SO_2-limestone slurry	7	8	9
SO_2-lime spray dryer	5	6	7
Particulate-ESP	4	5	6
Particulate-baghouse	5	6	7
Landfill (disposal only)	3	3	3

Plant and administrative overheads include plant services, general engineering (excluding maintenance), and the expenses connected with management activities. Plant and administrative overheads are 60% of the conversion costs less utilities.

Capital Charges--

Capital charges are those costs incurred by construction of the facility that must be recovered during its life. They consist of returns on equity and debt (discount rate), depreciation, income taxes, and other costs such as insurance and local taxes. In keeping with common practice for investor-owned utilities, the weighted cost of capital is used as the discount rate (121). Depreciation is stated as a sinking fund factor to simplify calculations. An allowance for interim replacement is included. Credits are also included for tax preference allowances. The capital charges are shown below.

	% of total capital investment
Weighted cost of capital	10.00
Depreciation (sinking fund factor)	0.61
Annual interim replacement	0.56
Levelized accelerated tax depreciation	-1.36
Levelized investment tax credit	-1.93
Levelized income tax	4.31
Insurance and property taxes	2.50
Levelized annual capital charge	14.7

The capital charges are applied as a percentage of the total capital investment.

The capital structure is assumed to be 35% common stock, 15% preferred stock, and 50% long-term debt. The cost of capital is assumed to be 11.4% for common stock, 10.0% for preferred stock, and 9.0% for long-term debt. The weighted cost of capital is 10.0%. The discount rate is equal to the weighted cost of capital. Other economic factors used in financial calculations are a 10% investment tax credit rate, 50% State plus Federal income taxes, 2.5% property tax and insurance, and an annual inflation rate of 6%. Salvage value is assumed to be less than 10% and equal to removal cost.

The sinking fund factor method of depreciation is used since it is equivalent to straight-line depreciation levelized for the economic life of the facility using the weighted cost of capital. The use of the sinking fund factor does not suggest that regulated utilities commonly use sinking fund depreciation. An annual interim replacement (retirement dispersion) allowance (122) is also included as an adjustment to the depreciation account to ensure that the initial investment will be recovered within the actual rather than the forecasted life of the facility. Tax preference allowances are incentives designed to encourage investment as a stimulus to the overall economy. The basic accounting method used is the flow-through method which passes the tax advantages to revenue requirements as soon as they occur.

Levelized Operating and Maintenance Costs--

Assuming a constant inflation rate, the levelized operating and maintenance costs are determined by multiplying the first-year operating and maintenance costs by an appropriate levelizing factor, L_f. The levelizing factor is calculated as follows:

$$L_f = CRF_e\ (K + K^2 + K^3 + --- + K^N)$$
$$= CRF_e\ [K(1 - K^N)]/(1 - K)$$

where: CRF_e = capital recovery factor for the economic life
$K = (1 + i)/(1 + r)$; present worth of an inflationary value
i = inflation rate
r = discount rate
N = book life in years

An inflation rate of 6% ($i = 0.06$) and a discount rate of 10% ($r = 0.10$) are used for new units. The first-year operating and maintenance costs are multiplied by the levelizing factor to obtain the levelized operating and maintenance costs.

ACCURACY OF ESTIMATES

The accuracy of the capital estimates used in this evaluation is -15%, +30%. It represents the potential variation in costs for an actual installation in comparison to the estimated costs, expressed as a percent of the estimated cost. The accuracy assigned to a cost estimate is empirical and not related to variabilities in a statistical sense. Rather, it depends on both the quantity and the quality of the technical data available. Accuracy ranges also reflect the numerous uncertainties surrounding estimates made using simplifying assumptions. However, when comparing the costs for processes evaluated using the same methodology, many of the same simplifying assumptions are made for each of the processes. Therefore, the comparability is greater than the overall accuracy of the estimates. When directly comparing estimates of the same grade, the uncertainty ranges associated with the compared costs are estimated to be less than ±10%.

SYSTEMS ESTIMATED

The base case systems for the three cases are described in this section. Each of the three processes for NO_x, SO_2, and particulate control in each case is defined by a flow diagram, a material balance, and a major equipment list (the costs shown in the equipment list are the installed costs exclusive of ancillary equipment and facilities such as foundations, structures, piping, electrical equipment, and control systems). The three processes are treated separately and are further divided into processing areas to facilitate cost comparisons.

NO_x Control	Case 1	Case 2	Case 3
Ammonia storage and injection	Area 1	Area 1	Area 1
Reactor	Area 2	Area 2	Area 2
Flue gas handling	Area 3	Area 3	Area 3
Air heater modification	Area 4	Area 4	Area 4
Waste disposal	Area 5	Area 5	Area 5
SO_2 Control			
Materials handling	Area 6	Area 6	Area 6
Feed preparation	Area 7	Area 7	Area 7
Flue gas handling	Area 8	Area 8	Area 8
SO_2 absorption	Area 9	Area 9	Area 9
Reheat	Area 10	None	None
Oxidation	Area 11	None	None
Lime particulate recycle	None	Area 10	None
Solids separation	Area 12	None	Area 10
Waste disposal	Area 13	Area 11	Area 11
Particulate Control			
Particulate removal and storage	Area 14	Area 12	Area 12
Particulate transfer	Area 15	Area 13	Area 13
Flue gas handling	Area 16	Area 14	Area 14
Waste disposal	Area 17	Area 15	Area 15

All of the cases are based on 500-MW power units and include, in addition to the three emission control processes, the costs for bottom ash collection, dewatering, and disposal. In most cases in which identical or similar functions such as flue gas handling and landfill are shared by two or three processes, the costs are prorated to each process.

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CASE 1

Case 1 is based on 3.5% eastern bituminous coal. It consists of an SCR NO_x control process, a limestone FGD process with forced oxidation, and cold-side ESPs. The flow diagram is shown in Figure 3, the material balance is shown in Table 16, and the equipment list is shown in Table 17. Each of the processing areas is described below.

NO_x Control

Processing areas 1 through 5 describe the NO_x control process.

Area 1 - Ammonia Storage and Injection--

A 30-day supply of liquid ammonia is stored in 250-psig storage tanks. A compressor is provided to unload the ammonia from truck or rail transporters (a spare is also provided). The tanks are equipped with water sprays to prevent overpressurization in hot weather and a water spray absorber system for purging and emergency discharge.

Electric heaters at the storage tanks heat the ammonia which discharges into an accumulator tank. The accumulator supplies two ammonia metering systems, one for each reactor train. The ammonia is metered at an $NH_3:NO_x$ mol ratio of 0.81 into an ammonia-air mixer where it is mixed with preheated 250°F air at a ratio of 1 part ammonia to 29 parts air. This dilutes the ammonia to a concentration outside the flammability range (15% to 27%) and improves the flow and mixing characteristics. The ammonia-air mixture is conveyed to the injection grids in insulated ducts.

The injection grid consists of a Cor-Ten pipe array in the reactor inlet duct. A Cor-Ten mixing grid downstream from the injection grid ensures uniform mixing of the ammonia-air mixture with the flue gas.

Area 2 - Reactor--

Two reactor trains are used. Each reactor consists of a carbon steel vessel 46 feet by 37 feet in size and 43 feet high, supported 40 feet above the level of the power plant floor. The reactors are equipped with ash hopper bottoms (costed in area 15) and are insulated with 6 inches of mineral wool. The flue gas enters at the top of the reactor and leaves at the bottom through a side duct above the ash hopper. The reactors are equipped with access doors and internal supports and structures for catalyst handling and support.

The catalyst is a honeycomb type supplied in blocks 150-mm square and 1 meter in length (the direction of flow). The blocks are mounted in metal baskets to form modules and placed in the reactor to form vertically separated beds. The catalyst volume is 24,400 ft^3, which provides a space velocity of 2,350 hr^{-1}. The catalyst modules are loaded into and removed from the

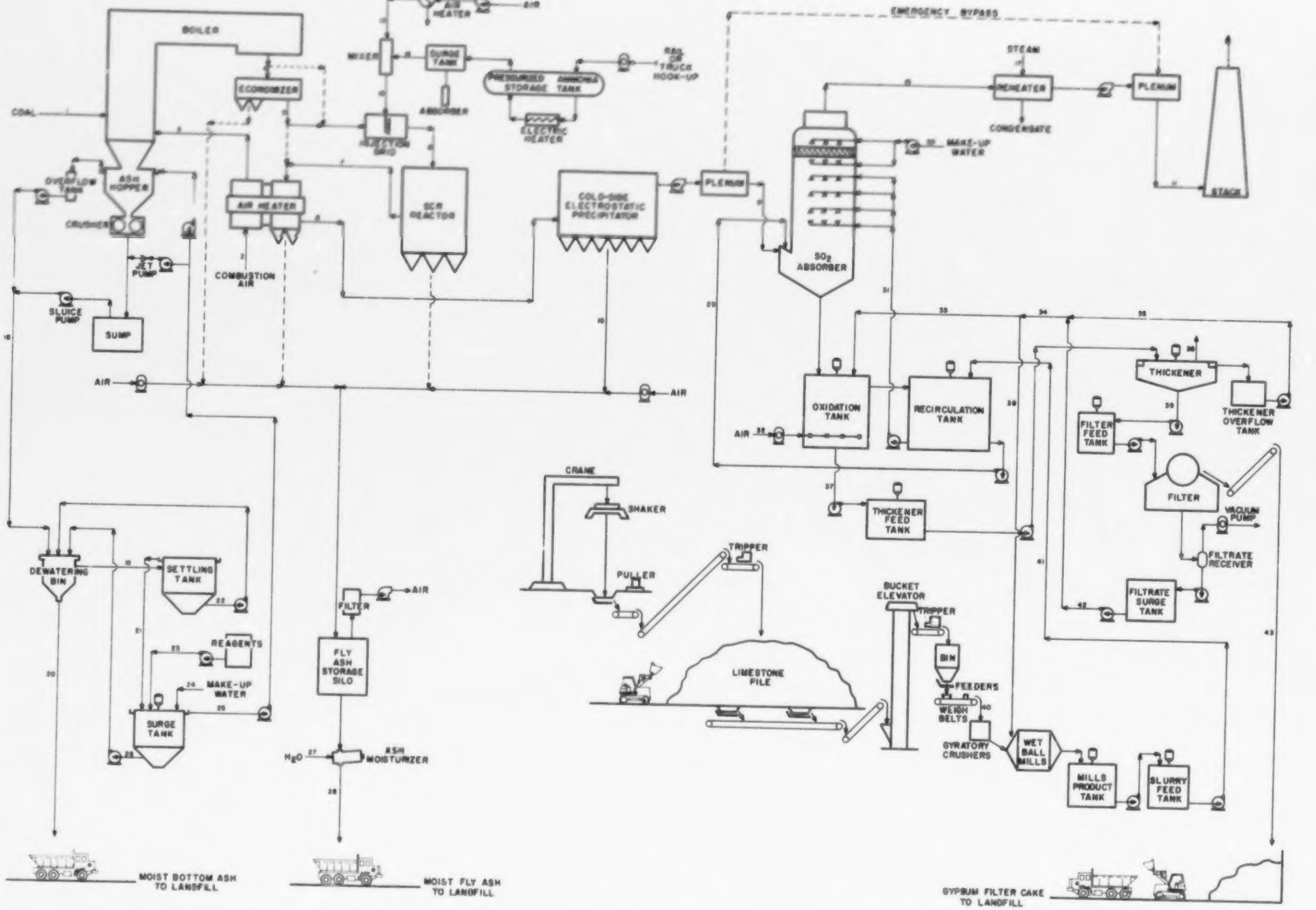
BOILER
COAL
ECONOMIZER
ASH HOPPER
OVERFLOW TANK
CRUSHER
JET PUMP
SLUICE PUMP
SUMP
AIR HEATER
COMBUSTION AIR
STEAM
STEAM AIR HEATER
AIR
MIXER
SURGE TANK
ABSORBER
INJECTION GRID
PRESSURIZED AMMONIA STORAGE TANK
ELECTRIC HEATER
RAIL OR TRUCK HOOK-UP
SCR REACTOR
COLD-SIDE ELECTROSTATIC PRECIPITATOR
PLENUM
EMERGENCY BYPASS
SO2 ABSORBER
MAKE-UP WATER
REHEATER
CONDENSATE
PLENUM
STACK
AIR
OXIDATION TANK
RECIRCULATION TANK
THICKENER FEED TANK
THICKENER
THICKENER OVERFLOW TANK
FILTER FEED TANK
FILTER
VACUUM PUMP
FILTRATE RECEIVER
FILTRATE SURGE TANK
CRANE
SHAKER
PULLER
TRIPPER
LIMESTONE PILE
BUCKET ELEVATOR
TRIPPER
BIN
FEEDERS
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
MILLS PRODUCT TANK
SLURRY FEED TANK
DEWATERING BIN
SETTLING TANK
REAGENTS
MAKE-UP WATER
SURGE TANK
FILTER
FLY ASH STORAGE SILO
H2O
ASH MOISTURIZER
MOIST BOTTOM ASH TO LANDFILL
MOIST FLY ASH TO LANDFILL
GYPSUM FILTER CAKE TO LANDFILL
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Figure 3. Case 1 flow diagram.

TABLE 16. CASE 1 MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to ammonia injection grid
1	Total stream, lb/hr	406,000	5,071,700	4,378,400	4,771,900	4,771,900
2						
3	Sft^3/min (60^{o}F)		1,120,700	967,500	1,003,100	1,003,100
4	Temperature, ^{o}F		80			705
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	6	7	8	9	10
	Description	Gas with ammonia to SCR reactor	Gas to air heater	Gas to ESP	Gas to spray tower	Gas from spray tower
1	Total stream, lb/hr	4,814,400	4,814,400	5,507,700	5,457,900	5,687,200
2						
3	Sft^3/min (60^{o}F)	1,012,600	1,012,600	1,165,800	1,165,800	1,249,200
4	Temperature, ^{o}F	701	705	300	300	125
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	11	12	13	14	15
	Description	Gas to stack	Steam to dilution air heater	Dilution air to mixer	Ammonia to mixer	Ammonia-air mixture to injection grid
1	Total stream, lb/hr	5,687,200	2,200	41,645	855	42,500
2						
3	Sft^3/min (60^{o}F)	1,249,200				
4	Temperature, ^{o}F	175	298	9,200	300	9,500
5	Pressure, psig		50	250		
6						
7						
8						
9						
10						

	Stream No.	16	17	18	19	20
	Description	Fly ash to storage silo	Steam to reheater	Bottom ash from boiler	Bottom ash sluice water to settling tank	Dewatered bottom ash to disposal
1	Total stream, lb/hr	49,800	99,500	74,900	92,700	13,900
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F	300	470			
5	Pressure, psig					
6						
7						
8						
9						
10						

(Continued)

TABLE 16. (Continued)

	Stream No.	21	22	23	24	25
	Description	Settling tank overflow to surge tank	Settling tank solids return to dewatering bin	Reagent to surge tank	Makeup water	Water to bottom ash sluice
1	Total stream, lb/hr	64,400	28,300	100	1,300	62,400
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	26	27	28	29	30
	Description	Surge tank underflow to dewatering bins	Mositurizer water	Fly ash to landfill	Recycle slurry to presaturator	Makeup water to spray tower
1	Total stream, lb/hr	3,400	5,500	55,300	2,946,200	288,300
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	31	32	33	34	35
	Description	Recycle slurry to spray tower	Air to oxidation tank	Clear liquid to oxidation tank	Clear liquid return	Thickener overflow return
1	Total stream, lb/hr	77,935,300	116,400	801,100	833,300	732,700
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F		25,700			
5	Pressure, psig		80			
6						
7						
8						
9						
10						

	Stream No.	36	37	38	39	40
	Description	Thickener evaporation	Slurry to thickener	Clear liquid to wet ball mills	Thickener bottoms to filter	Limestone to wet ball mills
1	Total stream, lb/hr	27,400	950,300	32,200	190,200	54,400
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F					
5	Pressure, psig					
6						
7						
8						
9						
10						

(Continued)

TABLE 16. (Continued)

Stream No.		41	42	43		
Description		Limestone slurry to recirculation tank	Filtrate return	Filter cake to landfill		
1	Total stream, lb/hr	86,600	100,600	89,600		
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F					
5	Pressure, psig					
6						
7						
8						
9						
10						

Stream No.						
Description						
1	Total stream, lb/hr					
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F					
5	Pressure, psig					
6						
7						
8						
9						
10						

Stream No.						
Description						
1	Total stream, lb/hr					
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F					
5	Pressure, psig					
6						
7						
8						
9						
10						

Stream No.						
Description						
1	Total stream, lb/hr					
2						
3	Sft^3/min (60^{o}F)					
4	Temperature, ^{o}F					
5	Pressure, psig					
6						
7						
8						
9						
10						

TABLE 17. CASE 1 EQUIPMENT LIST

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 1--Ammonia Storage and Injection		
1. Compressor, NH_3 unloading (2): 14.6 ft3/min capable of 250 psig suction max., 5-hp motor, cast iron body, insulated (1 operating, 1 spare)	8,600	2,100
2. Tank, NH_3 storage (5): Horizontal, 9-ft diameter x 66 ft long, 30,000 gal, 250 psig, carbon steel	169,500	3,900
3. Vaporizer, NH_3 (55): Electric resistance heater, carbon steel shell, 15-kW rated, 11 per ammonia storage tank	32,000	900
4. Tank, ammonia accumulator (1): 293 ft3, 5.5-ft diameter x 10.5 ft long, carbon steel, insulated (3 in.), +2.75-ft hemispherical end, 15 psig design pressure, 180°F design temperature	5,200	5,200
5. Ammonia absorber (1): 4 ft high x 1.1-ft diameter, 1.5-ft support, with vent, water supply, 1/4-in. Cor-Ten	400	1,900
6. Blower, air (3): 4,800 aft3/min, 20 in. H_2O, 25 hp, carbon steel, insulated (2 operating, 1 spare)	13,100	1,400
7. Heater, dilution air (2): Fin tube steam heater, 540-ft^2 surface area, aluminum tubes, galvanized cabinet	27,500	800
8. Mixer, ammonia and dilution air (2): 32-in. diameter x 10 ft long, carbon steel	12,400	7,800
9. Injection grid, NH_3 and air (2): 25 ft wide, 19 ft high, Cor-Ten pipe and supports	77,300	79,300
10. Mixing, grid; NH_3, air, and flue gas (2): 26 ft wide, 15 ft high, Cor-Ten pipe	9,500	24,600
Total, Area 1	355,500	127,900

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 2--Reactor		
1. Reactor (2): 46 ft wide x 37 ft long x 43 ft high, 6-in. mineral wool insulation; carbon steel housing, internals, and supports; elevated 40 ft	2,191,200	2,327,700
2. Sootblower, steam (20): 46 ft, retractable, 40-lb/min steam at 86 psig, 1 hp	520,000	33,100
3. Reactor crane and hoist (2): Electric 2-speed hoist, 2-ton capacity, 80-ft lift, grade to access door, 3 hp	21,200	600
4. Reactor hoist (4): Electric single-speed hoist, 2-ton capacity, access door to inside reactor, 3 hp	28,200	1,700
Total, Area 2	2,760,600	2,363,100
Area 3--Flue Gas Handling Modifications[a]		
1. Fan, flue gas (2): Induced draft, 862,243 aft^3/min, ΔP = 22 in. H_2O, carbon steel, 4,000 hp, fluid drive, double width, double inlet	73,100	300
Total, Area 3	73,100	300
Area 4--Air Heater Modifications[b]		
1. Air heater (2): Vertical inverted, size 31, Ljungstrom type,	575,000	10,300
Hot elements: DN type, 22 gauge, low alloy corrosion resistant, 16 in. deep, 84,900-ft^2 area		

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Cold elements: NF type, 3.5-mm spacing, 22 gauge, low alloy corrosion resistant, 42 in. deep, 274,500-ft^2 area		
2. Sootblower, steam (2): Retractable, 127-lb/min steam at 200 psig	15,200	900
3. Pump, wash water booster (3): Centrifugal, 2,020 gpm, 210-ft head, 200 hp, carbon steel, (2 operating, 1 spare)	3,550	
Total, Area 4	625,700	14,200
Area 5--Waste Disposal[c]		
1. Landfill site development and construction (1): 161-acre landfill site, 2,256-ft square landfill, 10,144,000 yd^3 volume, 30-yr life, 139 ft high at center, 9,171-ft perimeter ditch to 141,000-yd^3 catchment basin	12,900	1,200
2. Wheel loader (2): 7.0-yd^3 bucket, diesel engine	1,900	
3. Dozer (2): Track type with straight blade, 137-hp diesel engine	700	
4. Compactor (2): Vibratory sheepsfoot compactor, self-propelled	1,000	
5. Wheel loader (1): 3.5-yd^3 bucket, diesel engine	300	
6. Water truck (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	100	

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
7. Service truck (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	100	
8. Onsite trailer for sanitary facilities and break room (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply, and 120-volt electric supply	100	
9. Onsite water supply and discharge treatment system (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps	100	100
10. Truck (4): Tandem-axle, 4-rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesel engine (3 operating, 1 spare), 0.2% of total truck costs in this area	500	
Total, Area 5	17,700	1,300

a. Costs shown are additional costs of boiler I.D. fan due to NO_x reactor pressure loss.

b. Costs shown are for modifications and additional equipment made necessary by NO_x removal.

c. Except as noted, 0.3% of total waste disposal costs is charged to NO_x removal.

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 6--Materials Handling		
1. Car shaker and crane (1): Top mounting with crane, 20-hp shaker, 7-1/2-hp hoist	71,900	13,000
2. Car puller (1): 25-hp puller, 5-hp return	63,000	19,600
3. Hopper, limestone unloading (1): 16-ft diameter x 10-ft straight side, 2,650 ft3, carbon steel, 50-degree cone bottom, includes 6-in. square grating	15,500	5,900
4. Feeder, limestone unloading (1): Vibrating pan, 30 in. wide x 60 in. long, 3.5 hp, 250 ton/hr, carbon steel	5,500	500
5. Conveyor, limestone unloading (1): Belt 36 in. wide x 20 ft long, 5 hp, 250 ton/hr, 165 ft/min	11,400	1,400
6. Conveyor, limestone unloading (incline) (1): Belt, 36 in. wide x 310 ft long, 50 hp, 15-degree incline, 250 ton/hr, 165 ft/min	85,300	4,800
7. Dust collector, limestone unloading pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning, includes 5 dust hoods	11,200	5,200
8. Pump, limestone sump pit (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400	800
9. Conveyor, storage (1): Belt, 36 in. wide x 200 ft long, 5 hp, 250 ton/hr, 165 ft/min	73,100	3,900
10. Tripper, storage conveyor (1): 1 hp, 30 ft/min	27,200	9,100
11. Mobile, equipment (1): Scraper tractor, 3.0-yd3 bucket, 170-hp diesel engine	141,900	0
12. Hopper, reclaim (2): 7-ft square x 4-1/4 ft deep x 2-ft-wide bottom, 75 ft3, carbon steel, 60-degree cone bottom	2,400	1,600

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
13. Feeder, reclaim (2): Vibrating pan, 3.5 hp, 100 ton/hr	10,900	1,100
14. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning, 4 hoods	7,800	2,600
15. Pump, reclaim sump pit (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400	800
16. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 200 ft long, 5 hp, 100 ton/hr, 105 ft/min	40,900	2,900
17. Conveyor, limestone reclaim (incline) (1): Belt, 30 in. wide x 193 ft long, 40 hp, 15-degree incline, 50-ft lift, 100 ton/hr, 105 ft/min	60,300	3,700
18. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90-ft lift, 100 ton/hr	57,800	6,700
19. Conveyor, live limestone feed (1): Belt, 30 in. wide x 60 ft long, 7-1/2 hp, 100 ton/hr, 105 ft/min	20,500	1,400
20. Tripper, live limestone feed conveyor (1): 1 hp, 30 ft/min	27,200	9,100
21. Bin, crusher feed (3): 13-ft diameter x 21-ft straight side height, 3,100 ft3, covered, 50-degree cone, carbon steel	43,300	24,100
Total, Area 6	781,900	118,200
Area 7--Feed Preparation		
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp	49,600	2,300

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
2. Crusher (3): Gyratory, 75 hp	297,100	6,500
3. Ball mill, wet (3): Wet, open system, 9-ft diameter x 19 ft long, 741 hp, 13.0 ton/hr (2 operating, 1 spare)	1,721,900	122,000
4. Dust collector, ball mill (3): Bag filter, polypropylene bag, 2,200 aft3/min, 7.5 hp, reverse jet cleaning	23,300	7,800
5. Tank, mills product (3): 10-ft diameter x 10 ft high, 5,500 gal, open top, four 10-in. baffles, agitator supports, carbon steel, glass-filled polyester lining	13,700	11,000
6. Agitator, mills product tank (3): 40-in. diameter, 10 hp, neoprene coated	22,900	5,500
7. Pump, mills product tank (3): Centrifugal, 55 gpm, 60-ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,500	2,700
8. Tank, slurry feed (1): 21-ft diameter x 21 ft high, 57,800 gal, open top, four 20-in. baffles, agitator supports, carbon steel, glass-filled polyester lining	19,900	16,400
9. Agitator, slurry feed (1): 82-in. diameter, 50 hp, neoprene coated	43,000	3,500
10. Pump, slurry feed (8): Centrifugal, 27 gpm, 60-ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spares)	21,600	7,300
Total, Area 7	2,221,500	185,000

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 8--Flue Gas Handling		
1. Fans, flue gas (5): Induced draft, 380,800 aft^3/min, ΔP = 7.8 in. H_2O, 669 hp, fluid drive, double width, double inlet, Inconel 625	3,420,100	59,600
Total, Area 8	3,420,100	59,600
Area 9--SO_2 Absorption		
1. SO_2 absorber (5): Spray tower, 40 ft high x 34 ft wide x 17 ft deep, 1/4-in. carbon steel, neoprene lining, 316 stainless steel grids, FRP chevron vane entrainment separator, slurry header and nozzles	5,560,300	452,500
2. Tank, recirculation (5): 38-ft diameter x 38 ft high, 326,000 gal, open top, four 38-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining	360,500	291,200
3. Agitator, recirculation tank (5): 158-in. diameter, 67 hp, neoprene coated	445,400	182,700
4. Pump, presaturator (10): Centrifugal, 1,401 gpm, 100-ft head, 62 hp, carbon steel, neoprene lined (4 operating, 6 spares)	103,500	32,400
5. Pump, recirculation (15): Centrifugal, 18,525 gpm, 100-ft head, 819 hp, carbon steel, neoprene lined (8 operating, 7 spares)	1,697,900	150,200
6. Pump, makeup water (2): Centrifugal, 3,501 gpm, 200-ft head, 295 hp, carbon steel (1 operating, 1 spare)	33,400	3,800

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
7. Sootblowers (40): Air-fixed	112,000	104,000
Total, Area 9	8,313,000	1,216,800
Area 10--Reheat		
1. Reheater (5): Steam, tube type, 2,750 ft^2, one-half of tubes made of Inconel 625 and one-half made of Cor-Ten	2,656,900	168,800
2. Sootblower (20): Air-retractable	183,000	78,200
Total, Area 10	2,839,900	247,000
Area 11--Oxidation		
1. Tank, oxidation (5): 30.2-ft diameter x 38.1 ft high, 203,800 gal, open top, four 30-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining, includes air sparger	274,600	222,000
2. Agitator, oxidation tank (5): 110-in. diameter, 59 hp, neoprene coated	319,500	131,000
3. Pump, oxidation bleed (8): Centrifugal, 452 gpm, 60-ft head, 12 hp, carbon steel, neoprene lined (4 operating, 4 spares)	37,700	15,800
4. Blower, oxidation air (6): 3,300 sft^3/min, 323 hp (4 operating, 2 spares)	208,600	4,700
5. Oxidation sparger (5): 19.1-ft-diameter ring	93,800	41,600
Total, Area 11	934,200	415,100

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 12--Solids Separation		
1. Tank, thickener feed (1): 19.1-ft diameter x 38.1 ft high, 81,400 gal, open top, four 20-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining	28,700	23,700
2. Agitator, thickener feed tank (1): 80-in. diameter, 42 hp, neoprene coated	31,500	2,600
3. Pump, thickener feed (2): Centrifugal, 1,807 gpm, 60-ft head, 48 hp, carbon steel, neoprene lined (1 operating, 1 spare)	19,200	7,200
4. Thickener (1): 48-ft diameter x 5.4 ft high, carbon steel sides, concrete basin, includes 1-hp rake motor and mechanism, 1,780-ft^2 area	85,500	59,300
5. Tank, thickener overflow (1): 27.5-ft diameter x 5.4 ft high, 24,200 gal, open top, carbon steel	6,600	4,500
6. Pump, thickener overflow tank (2): Centrifugal, 1,466 gpm, 75-ft head, 46 hp, carbon steel, neoprene lined (1 operating, 1 spare)	12,900	1,500
7. Pump, thickener underflow (2): Centrifugal, 287 gpm, 9.3-ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,800	3,200
8. Tank, filter feed (1): 9.3-ft diameter x 9.3 ft high, 4,730 gal, open top, four 9-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining	3,700	3,100
9. Agitator, filter feed tank (1): 36-in. diameter, 7 hp, neoprene coated	5,600	500

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
10. Pump, filter feed tank (3): Centrifugal, 143 gpm, 50-ft head, 4 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,900	3,600
11. Filter (3): Rotary vacuum, 8-ft diameter x 14-ft face, 48 hp, includes auxiliary equipment, 380-ft^2 area (2 operating, 1 spare)	381,100	68,600
12. Pump, filtrate (4): Centrifugal, 101 gpm, 20-ft head, 1 hp, carbon steel, neoprene lined (2 operating, 2 spares)	17,300	1,900
13. Tank, filtrate surge (1): 8.3-ft diameter x 8.3 ft high, 3,300 gal, open top, carbon steel	1,700	1,200
14. Pump, filtrate surge tank (2): Centrifugal, 201 gpm, 85-ft head, 7 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,300	1,000
15. Conveyor, filtrate cake (1): Belt, 30 in. wide, 75-ft-long horizontal, 1-1/2 hp, 50 ton/hr, 100-ft incline	37,100	3,500
Total, Area 12	659,900	185,400
Area 13--Waste Disposal[a]		
1. Landfill site development and construction (1): 161-acre landfill site, 2,256-ft square landfill, 10,144,000-yd^3 volume, 30-yr life, 139 ft high at center, 9,171 ft perimeter ditch to 141,000-yd^3 catchment basin	2,616,300	243,800
2. Wheel loader (2): 7.0-yd^3 bucket, diesel engine	384,000	
3. Dozers (2): Track type with straight blade, 137-hp diesel engine	138,200	

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
4. **Compactor** (2): Vibratory sheepsfoot compactor, self-propelled		
5. **Wheel loader** (1): 3.5-yd3 bucket, diesel engine	69,000	
6. **Water truck** (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	17,700	
7. **Service truck** (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	45,100	
8. **Onsite trailer for sanitary facilities and break room** (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply, and 120-volt electric supply	3,700	
9. **Onsite water supply and discharge treatment system** (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps	27,500	22,500
10. **Trucks** (4): Tandem-axle, 4-rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesel engine (3 operating, 1 spare), 54.0% of total truck costs in this area	143,200	
Total, Area 13	3,648,600	266,300

a. Except as noted, 54.4% of total waste disposal costs is charged to SO_2 removal.

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 14--Particulate Removal and Storage		
1. Electrostatic precipitator, cold side (2): 849,850 aft3/min, 424,925-ft^2 collection area, 1.0-in. pressure drop, 99.72% removal efficiency, 500-ft^2/kaft3/min SCA, 48.5 ft deep x 71.8 ft wide x 38.8 ft high, (inside dimensions)	2,646,400	2,384,800
2. Hopper, economizer ash (8): Inverted pyramid-type double-V hopper, 15 ft long x 7.5 ft wide x 7.2 ft deep, thermally isolated design, constructed of 3/8-in. Cor-Ten plate, 55-degree valley angle, each hopper has 2 outlets, 216-ft3 volume and 244-ft^2 area per hopper	146,500	92,200
3. Hopper, air heater ash (8): Inverted pyramid-type double-V hopper, 15 ft long x 7.5 ft wide x 7.2 ft deep, constructed of 3/8-in. Cor-Ten plate, heat traced and insulated, 55-degree valley angle, each hopper has 2 outlets, 216-ft3 volume and 244-ft^2 area per hopper, 6-kW heater	77,600	45,100
4. Hopper, ESP ash (20): Inverted pyramid-type double-V hopper, 24.3 ft long x 14.4 ft wide x 13.8 ft deep, constructed of 3/8-in. Cor-Ten plate, heat traced and insulated, each hopper has 2 outlets, 55-degree valley angle, 1,430-ft3 volume and 802-ft^2 area per hopper, 10-kW heater	556,700	311,800
5. Hopper, NO_x reactor ash (12): Inverted pyramid-type double-V hopper, 23 ft long x 12.3 ft wide x 11.8 ft deep, constructed of 3/8-in. Cor-Ten plate, insulated, 55-degree valley angle, each hopper has 2 outlets, 949-ft3 volume and 634-ft^2 area per hopper, 10-kW heater	264,500	149,400

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
6. Hopper, bottom ash (1): 51 ft long x 9 ft wide x 22 ft high (inside dimensions), double-V hopper, center discharge with 3,320-ft3 capacity for 12-hr ash containment, supported independently of furnace-boiler, 3/8-in. carbon steel plate, refractory lined, 4 hydraulically operated exit doors emptying to 4 double-roll clinker grinders, 10-in. diameter x 2-ft-long manganese steel rolls, 60 hp	352,000	202,600
Total, Area 14	4,043,700	3,185,900
Area 15--Particulate Transfer		
1. Pressure pneumatic transfer system for fly ash (1):		
a. Conveying lines, pressure pneumatic for fly ashes (1): Pipelines and pipe fittings for pressure pneumatic conveyance of ash, 50-ton/hr conveying capacity with 1,320-ft equivalent length system, 10-in. I.D. branch lines and 12-in. I.D. main lines, nickel-chromium cast iron pipe with Ni-Hard or equivalent pipe fittings	59,800	32,000
b. Pressure feeders, ash and air (96): Materials-handling valve, electrically actuated, air operated, 10-in. I.D. ash inlet, 10-in. I.D. ash outlet, cast iron body, stainless steel slide gate; each assembly includes two spring-loaded, air-inlet check valves with cast iron bodies	768,000	437,200
c. Valves, line segregating (12): Segregating slide valve, electrically actuated, air operated for on-off control of each branch conveying line, 12-in. I.D. port, cast iron body, stainless steel slide gate	28,800	16,600

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
d. System control unit (1): Automatic sequence control unit to control the programmed operation of materials-handling valves, line segregating valves, and blowers; includes gauges for manual reading and override switches for manual operation	96,000	54,600
e. Filters, silo bag (2): Automatic cycling vent filter, 1,440-ft^2 bag area, 12 ft x 5.3 ft x 11 ft overall dimensions	40,000	23,200
f. Fans, bag filter vent (2): 8,000 aft^3/min, $\Delta P = 6$ in. H_2O, 20 hp	16,000	9,000
g. Compressor, pressure pneumatic transfer system (3): 4,525 aft^3/min, 13.75 psig, 500 hp, carbon steel, with silencers (2 operating, 1 spare)	108,000	9,300
h. Silo, fly ash storage (2): 30-ft diameter x 55 ft high, 35,300-ft volume, with bin air fluidizing system, elevated construction for 11-1/2-ft truck clearance, rotary star feeders and moisturizers, carbon steel plate, 20 hp	584,000	332,800
2. Bottom ash sluice transfer system (1):		
a. Pumps, bottom ash water supply (3): Centrifugal, 255 gpm, 90-ft head, carbon steel, 10 hp (2 operating, 1 spare)	10,400	1,800
b. Pumps, bottom ash water supply (3): Centrifugal, 1,860 gpm, 115-ft head, carbon steel, 75 hp (2 operating, 1 spare)	25,900	4,000
c. Pumps, bottom ash water supply (3): Centrifugal, 587 gpm, 577-ft head, carbon steel, 150 hp (2 operating, 1 spare)	40,600	5,200

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
d. Tank, overflow (1): 18-ft diameter x 8 ft high, 11,400 gal, flat bottom, open top, with an overflow weir 2 ft below top of tank, 3/8-in. carbon steel, epoxide-coated interior	11,400	11,000
e. Pump, bottom ash hopper overflow bin (3): Centrifugal, 550 gpm, 175-ft head, carbon steel, 40 hp (2 operating, 1 spare)	17,400	2,700
f. Jet pump, bottom ash conveyance (4): Jet ejector nozzle assembly and adapter to bottom ash hopper, 360 gpm, 692-ft head supply water, Ni-Hard nozzle and throat construction (2 operating, 2 spares)	4,000	1,600
g. Sump pit, sluice (1): Concrete pit, 5 ft wide x 5 ft long x 8 ft deep with two agitator nozzles located in bottom of bin to prevent settling	2,900	6,300
h. Pumps, bottom ash sluice (3): Centrifugal, slurry pumps, 2,550 gpm, 230-ft head, Ni-Hard liner and impeller, 250 hp (2 operating, 1 spare)	108,100	6,200
i. Valves, shutoff and crossover (17): Air-operated gate valve, 8-in. I.D. port, Ni-Hard	30,400	17,400
j. Slurry pipeline, one-quarter mile basalt-lined to dewatering bins, normal use (1): Pipeline comprising 74, 18-ft-long sections of flanged, basalt-lined steel pipe, 8-in. I.D. and 4 basalt-lined elbows or bends, 8-in. I.D.	82,000	29,500

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
k. Slurry pipeline, spare line to dewatering bins and return waterline (2): Pipeline comprising 34, 40-ft-long sections of flanged steel pipe, 8-in. I.D., schedule 80 carbon steel, and 4 hardened elbows or bends, 8-in. I.D.	31,200	11,000
l. Bin, bottom ash dewatering (2): Conical-bottom dewatering bin, 35-ft diameter x 64 ft high, with 18-1/2-ft cylindrical section, 26-ft-high cone, 17,160-ft3 volume, stainless steel floating decanter and movable drainpipe, stationary decanter in conical section, erected for 16-1/2-ft truck clearance, carbon steel with stainless steel decanter drums, 400-ton capacity	240,000	111,800
m. Settling tank, bottom ash return water (1): 50-ft diameter x 15 ft deep, 220,700 gal, carbon steel, epoxide-coated interior, open top	82,400	38,900
n. Surge tank, bottom ash return water (1): Water reservoir, 40-ft diameter x 16 ft deep, 154,100 gal, carbon steel, epoxide-coated interior, open top	62,400	29,000
o. Pump, underflow solids recycle (3): Centrifugal, 250 gpm, 100-ft head, Ni-Hard steel body and impeller, 15 hp (2 operating, 1 spare)	11,300	2,300
p. Pump, dewatering bin sump pit (3): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined (2 operating, 1 spare)	7,200	2,500
3. Water treatment system for recycle water alkalinity control (1):		

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
a. Storage tank, sulfuric acid for pH control of water (1): Cylindrical steel tank, 5-ft, 7-in. diameter x 5 ft, 7 in. high, 1,000 gal, flat bottom and closed flat top, carbon steel, all-weather housing	1,900	300
b. Metering pump, sulfuric acid (2): Positive displacement metering pump, 0.01 to 1 gpm, 0 psig, with flow rate controlled by a pH controller, Carpenter 20 or alloy of similar corrosion resistance to 93% sulfuric acid, 0.25 hp (1 operating, 1 spare)	1,900	600
c. Agitator, treated water (1): Agitator with 24-in.-diameter nickel-chromium blade, 5 hp	2,900	400
Total, Area 15	2,474,900	1,197,200
Area 16--Flue Gas-Handling Modifications[a]		
1. Fan, flue gas (2): Induced draft, 862,243 aft3/min, ΔP = 22 in. H_2O, carbon steel, 4,000-hp motor, fluid drive, double width, double inlet	20,900	100
Total, Area 16	20,900	100
Area 17--Waste Disposal[b]		
1. Landfill site development and construction (1): 161-acre landfill site, 2,256-ft square landfill, 10,144,000-yd3 volume, 30-yr life, 139 ft high at center, 9,171-ft perimeter ditch to 141,000-yd3 catchment basin	2,179,700	203,100

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
2. Wheel loader (2): 7.0-yd3 bucket, diesel engine	319,900	
3. Dozers (2): Track type with straight blade, 137-hp diesel engine	115,200	
4. Compactor (2): Vibratory sheepsfoot compactor, self-propelled	169,900	
5. Wheel loader (1): 3.5-yd3 bucket, diesel engine	57,500	
6. Water truck (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	14,700	
7. Service truck (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	37,600	
8. Onsite trailer for sanitary facilities and break room (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply, and 120-volt electric supply	3,100	
9. Onsite water supply and discharge treatment system (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps	22,900	18,800

(Continued)

TABLE 17. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
10. Trucks (4): Tandem-axle, 4-rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesel engine (3 operating, 1 spare), 45.8% of total truck costs in this area	121,300	
Total, Area 17	3,041,800	221,900

a. Costs shown as additional costs of boiler I.D. fan due to ESP pressure loss.
b. Except as noted, 45.3% of total waste disposal costs is charged to ash removal.

reactor with two hoists; one lifts the modules from ground level and the other moves the modules in the reactor on a monorail system. Retractable steam sootblowers are provided between each catalyst bed to remove ash and ammonia salt deposits.

Area 3 - Flue Gas Handling--

This area includes the incremental increase in the capacity of the two boiler ID fans to compensate for the 7-inch H_2O pressure drop in the reactors and the ductwork associated with the NO_x control system. The ductwork consists of the economizer bypass, reactor bypass, ducts from the boiler to the reactors and from the reactors to the air heater, and the additional ductwork needed to connect the air heater to the downstream equipment (made necessary to accommodate the NO_x control system). Included in the cost of the ductwork are the costs of insulation, flanges, dampers, and expansion joints.

Area 4 - Air Heater Modifications--

Modifications to two air heaters to reduce the adverse effects of ammonia salt deposition and corrosion are provided. Only the incremental costs for these modifications, as compared with a standard air heater, are included. The hot- and cold-end elements are increased in thickness (from 24 to 22 gauge and 22 to 18 gauge, respectively) and a low-alloy corrosion-resistant metal is used for the hot-end element. The hot-end element depth is decreased and the cold-end element depth is increased, with a net increase in overall depth. The cold-end element spacing is reduced from 6 mm to 3.5 mm and the diameter of the rotor is increased. The overall result of these changes is an increase in heat transfer area of 50%, compared with a standard air heater.

Two additional steam sootblowers are provided for each air heater to clean the hot-end elements. The frequency of water washing and the quantity of water used are also increased, with a corresponding increase in the size of pipes and pumps in the water washing system. The increase in the quantities of steam and water used as a result of these changes is shown in Table 18.

Area 5 - Waste Disposal--

The only disposal costs involved in NO_x control are for spent catalyst disposal. The catalyst is assumed to be a nonhazardous waste that is trucked to and disposed of in the landfill. The disposal costs are prorated from the total waste disposal costs based on volume.

SO_2 Control

Processing areas 6 through 13 describe the limestone FGD process.

Area 6 - Materials Handling--

The materials-handling area comprises the equipment to unload, store, and transfer the limestone used in the FGD system. The 0- x 1-1/2-inch limestone is dumped from trucks or railcars to an unloading hopper and transported by

TABLE 18. STEAM SOOTBLOWING AND WATER WASHING REQUIREMENTS FOR AIR HEATERS OF CASE 1

	Steam sootblowing					
Case	Number of blowers	Cycles/day/ blower	Min/cycle	Lb steam/min	Lb steam/yr	Additional lb steam/yr
Standard without SCR	2	3	17	127	2,968,625	
Modified with SCR	4	3	18	127	6,286,500	3,317,875

	Water washing					
Case	Cycles/yr	Hr/cycle	Gal/ min/heater	Psig	Gal/yr	Additional gal/yr
Standard without SCR	1	8	1,320	75	1,267,200	
Modified with SCR	2	8	2,020	150	3,878,400	2,611,200

belt conveyors to a 30-day capacity storage pile over two reclaim hoppers. The unloading system has a capacity of 250 ton/hr. The limestone is reclaimed and transported to the feed preparation area with a belt conveyor system with a capacity of 100 ton/hr. The system is operated intermittently to meet the FGD limestone requirements of 27 ton/hr.

Area 7 - Feed Preparation--

The feed preparation area consists of equipment to prepare limestone slurry for the FGD system. It includes two operating and one spare train, each consisting of a crusher, a wet ball mill, and an agitated product tank, with ancillary equipment such as a dust collection system and pumps. The limestone is first crushed to 0 x 3/4 inch and then wet ground to a 60% slurry with a particle size of 90% minus 325 mesh. The slurry from each ball mill is collected in the mill product tank of the ball mill, then stored in an 8-hour capacity feed tank that supplies the FGD system.

Area 8 - Flue Gas Handling--

This area consists of the inlet plenum that supplies the absorber trains, the absorber train ductwork, two emergency bypass ducts from each end of the feed plenum to the stack plenum for bypass of 50% of the scrubbed gas, and one ID booster fan for each of the five absorber trains to compensate for the 7.8-inch H_2O pressure drop in the FGD system.

The emergency bypass ducts and the ductwork upstream from the absorbers and downstream from the reheaters are constructed of Cor-Ten steel. The ductwork from the absorbers to the reheaters is constructed of 316 stainless steel. The ID booster fans are constructed of Inconel 625. The flue gas velocity is 50 ft/sec. All ducts are insulated with 2 inches of glass wool.

Area 9 - SO_2 Absorption--

The area consists of the four operating and one spare spray tower absorbers and related equipment. The absorbers are equipped with presaturator systems in the inlet duct that spray the flue gas with 4 gal/kaft^3 of scrubbing liquid, cooling it from 300°F to 127°F as it enters the absorber. The spray towers are rectangular neoprene-lined carbon steel vessels 34 by 17 feet, 40 feet high with three layers of stainless steel grids to control the gas distribution. Each absorber contains four banks of spray headers, one above each grid spraying downward and one below the bottom grid spraying upward. The absorbers are equipped with horizontal open-vane, three-pass, fiberglass chevron mist eliminators to reduce the entrained moisture content of the flue gas to 0.1% by weight. The mist eliminators are continuously washed on the underside and intermittently on the top side with makeup water. The presaturators and absorbers are equipped with air sootblowers to remove deposits.

Absorbent liquid drains from the absorber into an oxidation tank (which is included in area 11) and overflows by gravity into a recirculation tank to which the makeup slurry is added. The absorbent liquid is recirculated from this tank to the presaturator and absorber spray headers.

The absorber is designed for a superficial gas velocity of 10 ft/sec and an SO_2 removal efficiency of 89% (in addition, 50% of the SO_3 and 100% of the HCl are removed). The L/G ratio is 106 gal/kaft^3 and the limestone stoichiometry is 1.4 mols $CaCO_3$/mol SO_2 plus 2HCl removed.

Area 10 - Reheat--

The reheaters are tubular steam heat exchangers designed to provide a flue gas temperature of 175°F at the stack plenum. They are situated in the ducts between the absorber and ID booster fan. The gas velocity is 25 ft/sec. The temperature increase required is approximately 47°F, from 125°F to 172°F, with the remaining increase provided by compression in the ID booster fans. The reheater tubes in contact with flue gas below 150°F are Inconel 625 and the remainder are Cor-Ten steel. The reheaters are equipped with air sootblowers to clean the tubes.

Area 11 - Oxidation--

The oxidation area for the FGD system consists of an agitated tank beneath each absorber that contains a sparging ring (a circular pipe manifold with holes in the periphery) to introduce oxidizing air and low-pressure compressors that supply air at a rate of 2.5-gm atoms O/gm mol SO_2 absorbed. The system is designed to produce a minimum oxidation level of 95%. A bleedstream containing 8% solids is withdrawn from the tank and pumped to the solids separation area; the remaining slurry overflows to the absorber recirculation tank.

Area 12 - Solids Separation--

In this area, the bleedstreams from the absorbers are dewatered and stored for removal to the disposal area. The 8% solids bleedstreams from the four absorber trains are combined in a thickener feed tank and dewatered to 40% solids in a 48-foot-diameter thickener. The thickener underflow is dewatered to 85% solids in two rotary vacuum filters (a spare filter is provided) and conveyed to a concrete storage pad. The filters are 8 feet in diameter and 14 feet long. The thickener overflow and the filtrate are returned to the feed preparation and absorber areas.

Area 13 - Waste Disposal--

The operations in this area consist of trucking the FGD waste to the common landfill and operation of the landfill. The waste is loaded into 26-yd^3 dump trucks with a front loader and hauled one mile to the landfill. The landfill and its operation are described in the premise section. The FGD waste disposal costs associated with the operation of the landfill are prorated on the basis of volume.

Particulate Control

Processing areas 14 through 17 describe the bottom ash and fly ash control processes.

Area 14 - Particulate Removal and Storage--

This area consists of two cold-side ESPs and all hoppers associated with bottom ash and fly ash collection. The ESPs are operated in parallel. Each is 49 feet long, 72 feet wide, and 39 feet high, with an SCA of 500 $ft^2/aft^3/min$ and a pressure drop of 1.0 inch H_2O. The removal efficiency is 99.7%. (Some reviewers state that an SCA range of 200 to 250 $ft^2/kaft^3/min$ is adequate to meet the ash removal required by the ESP in case 1. An SCA of 500 $ft^2/kaft^3/min$ was used for case 1 after determining SCA values ranging from about 450 to over 650 $ft^2/kaft^3/min$ from several references.)

There are 48 double-vee fly ash hoppers, each with 2 outlets, 8 on the economizer, 4 on each air heater, 10 on each ESP, and 6 on each SCR reactor. All are inverted-pyramid type with 55-degree slopes and are constructed of Cor-Ten steel. They are equipped with electric heaters and are insulated. The economizer hoppers are equipped with isolation chutes to prevent ash fusion and combustion of residual carbon. The bottom ash hopper has a double-vee bottom and a refractory lining. It is equipped with flushing jets and water lances and has four discharge doors, each with a 2-roll clinker grinder. All hoppers have a 12-hour capacity.

Area 15 - Particulate Transfer--

This area consists of a pneumatic system that removes the fly ash from the hoppers and silos where it is stored for transport to the landfill, and a bottom ash hydraulic transporting and dewatering system.

The fly ash pneumatic system operates at an air pressure of about 13 psig. Fly ash is removed from each of the 96 hopper outlets through air lock valves to 10-inch branch lines in an automatically controlled sequence. The branch lines connect to 12-inch main lines that transport the ash to two carbon steel storage silos, each 30 feet in diameter and 55 feet high, elevated for direct loading to trucks. The silos are equipped with fabric filter dust collectors, air fluidizing systems, and moisturizers to moisten the ash as it is discharged.

The bottom ash system consists of a hydraulic sluicing system to transport the ash to dewatering bins. The ash is periodically sluiced through the clinker grinders on the hopper to high-pressure water ejector pumps that sluice it to a sump from which it is sluiced one-fourth of a mile to the dewatering system by centrifugal pumps (the ejector pumps are self-priming and nonplugging while the centrifugal pumps are more efficient for long-distance pumping). All pipes are basalt lined and the pumps and fittings are constructed of abrasion-resistant metal alloy. The ash is sluiced to one of two dewatering bins. The bins have conical bottoms and are elevated for discharge of the ash to trucks. The bins have a capacity of 72 hours and operate on a 24-hour cycle to allow intermittent operation. The ash is allowed to drain to a 10% water content. The water drains first to a settling tank to remove fines and then to a surge tank. The water is returned to the sluicing system after pH adjustment and is also recirculated continuously through the bottom ash hopper as necessary to maintain a maximum 175°F water temperature in the hopper.

Area 16 - Flue Gas Handling--

This area includes the incremental increase in the boiler ID fan necessary to compensate for the 2-inch H_2O pressure drop in the ESP and related ductwork and the ductwork connecting the ESP to the air heaters and the FGD inlet plenum.

Area 17 - Waste Disposal--
This area consists of the equipment and operations involved in trucking the bottom ash and fly ash to the landfill and the portion of the landfill operation prorated to ash disposal.

CASE 2

Case 2 is based on 0.7% sulfur western subbituminous coal. It consists of an SCR NO_x control process, a lime spray dryer FGD process, and a baghouse for combined FGD and fly ash collection. The flow diagram is shown in Figure 4, the material balance is shown in Table 19, and the equipment list is shown in Table 20. Each of the process areas is described below with the exception of those for which the verbal description is similar to those for the same areas previously described in case 1.

NO_x Control

The 2-train SCR reactor system is essentially the same as the system in case 1. The major differences result from size differences related to the lower concentration of NO_x in the flue gas, the larger volume of flue gas, and slight differences in equipment layout resulting from these differences. Process areas 1 through 5 describe the NO_x control process.

Area 1 - Ammonia Storage and Handling--
The description of this area is the same as the description for area 1 in case 1. The ammonia storage system is identical. The ammonia vaporization and air heating and mixing system is slightly smaller because of the lesser quantity of ammonia required and the injection and mixing grids are slightly larger because of the larger volume of flue gas.

Area 2 - Reactor--
The equipment in this area is essentially the same as the equipment description for area 2 in case 1. The reactors are proportionally larger because of the larger flue gas volume (53 by 37 feet in cross section, compared with 46 by 37 feet in case 1) and the catalyst volume is 29,862 ft^3, providing a space velocity of 2,320 hr^{-1} (as compared with a catalyst volume of 24,400 ft^3 and a space velocity of 2,350 hr^{-1} in case 1).

Area 3 - Flue Gas Handling--
The pressure drop is 7 inches H_2O, as it is in case 1, and the ductwork design is the same but the equipment is proportionally larger because of the larger flue gas volume.

Area 4 - Air Heater Modifications--
The same air heater modifications described in case 1 are included. In this case, the air heater is more than double the size of that of a comparative air heater for service without NO_x control. The additional sootblowing and water washing requirements are shown in Table 21.

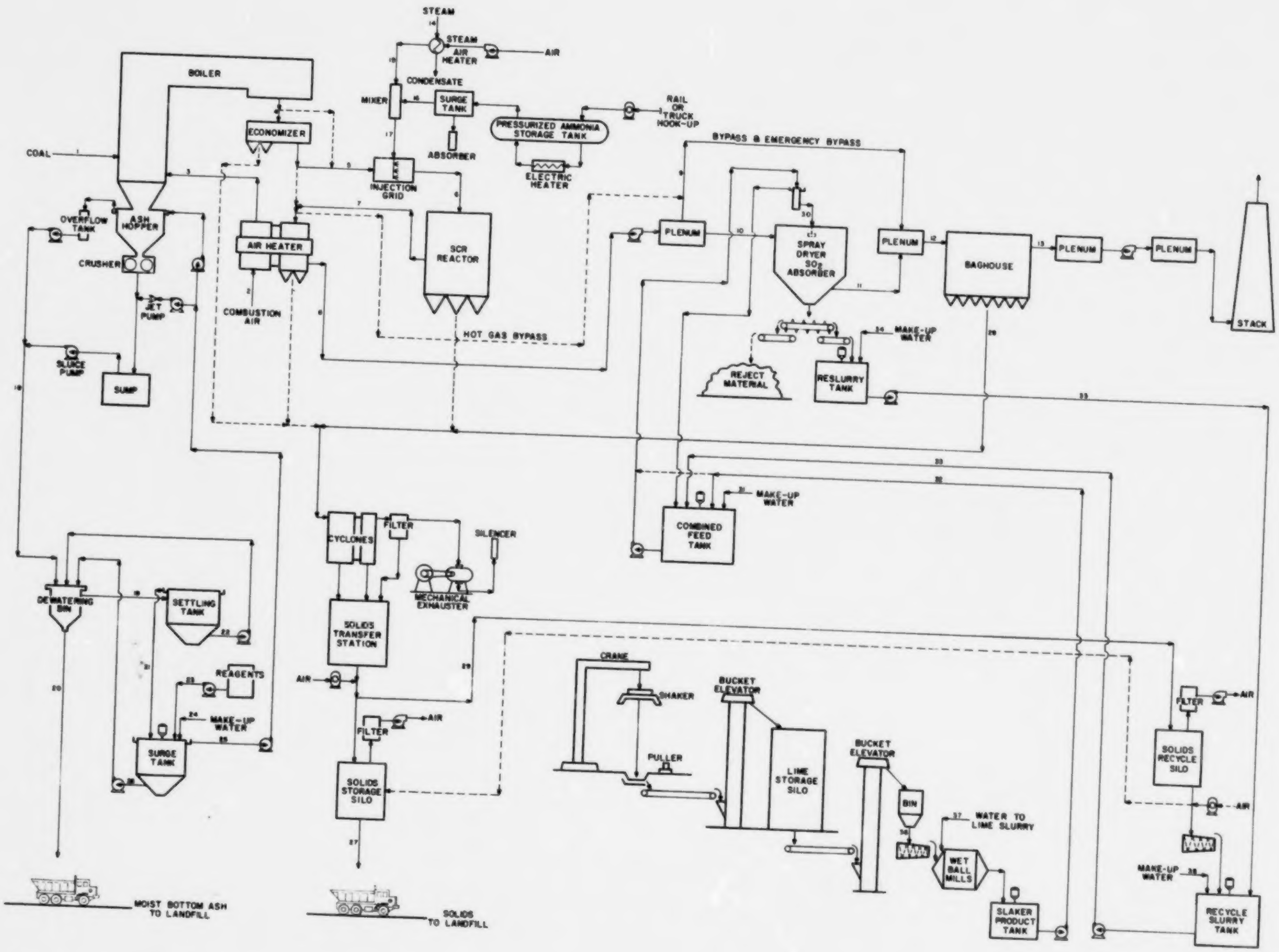
STEAM
STEAM AIR HEATER
AIR
CONDENSATE
MIXER
SURGE TANK
ABSORBER
PRESSURIZED AMMONIA STORAGE TANK
ELECTRIC HEATER
RAIL OR TRUCK HOOK-UP
BOILER
COAL
ECONOMIZER
INJECTION GRID
ASH HOPPER
OVERFLOW TANK
CRUSHER
JET PUMP
SLUICE PUMP
SUMP
AIR HEATER
COMBUSTION AIR
SCR REACTOR
HOT GAS BYPASS
BYPASS & EMERGENCY BYPASS
PLENUM
SPRAY DRYER SO2 ABSORBER
PLENUM
BAGHOUSE
PLENUM
PLENUM
STACK
REJECT MATERIAL
RESLURRY TANK
MAKE-UP WATER
COMBINED FEED TANK
MAKE-UP WATER
CYCLONES
FILTER
SILENCER
MECHANICAL EXHAUSTER
SOLIDS TRANSFER STATION
AIR
FILTER
AIR
SOLIDS STORAGE SILO
DEWATERING BIN
SETTLING TANK
REAGENTS
MAKE-UP WATER
SURGE TANK
MOIST BOTTOM ASH TO LANDFILL
SOLIDS TO LANDFILL
CRANE
SHAKER
PULLER
BUCKET ELEVATOR
LIME STORAGE SILO
BUCKET ELEVATOR
BIN
WATER TO LIME SLURRY
WET BALL MILLS
SLAKER PRODUCT TANK
FILTER
AIR
SOLIDS RECYCLE SILO
AIR
MAKE-UP WATER
RECYCLE SLURRY TANK

Figure 4. Case 2 flow diagram.

TABLE 19. CASE 2 MATERIAL BALANCE

Stream No.		1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to ammonia injection grid
1	Total stream, lb/hr	640,200	5,765,200	4,977,200	5,609,200	5,609,200
2						
3	Sft^3/min (60°F)		1,273,900	1,099,800	1,215,300	1,215,300
4	Temperature, $^\circ$F		80			750
5	Pressure, psig					
6						
7						
8						
9						
10						

Stream No.		6	7	8	9	10
	Description	Gas with ammonia to SCR reactor	Gas to air heater	Gas to inlet plenum	Spray dryer bypass gas	Gas to spray dryer
1	Total stream, lb/hr	5,648,400	5,648,400	6,436,400	742,200	5,694,200
2						
3	Sft^3/min (60°F)	1,224,100	1,224,200	1,398,300	161,200	1,237,100
4	Temperature, $^\circ$F	747	750	300	300	300
5	Pressure, psig					
6						
7						
8						
9						
10						

Stream No.		11	12	13	14	15
	Description	Gas from spray dryer	Combined gas to baghouse	Gas to stack	Steam to dilution air heater	Dilution air to mixer
1	Total stream, lb/hr	5,933,300	6,675,500	6,573,800	2,000	38,412
2						
3	Sft^3/min (60°F)	1,297,700	1,458,900	1,458,900		8,500
4	Temperature, $^\circ$F	154	170	170	298	250
5	Pressure, psig				50	
6						
7						
8						
9						
10						

Stream No.		16	17	18	19	20
	Description	Ammonia to mixer	Ammonia-air mixture to injection grid	Bottom ash from boiler	Bottom ash sluice water to settling tank	Dewatered bottom ash to disposal
1	Total stream, lb/hr	788	39,200	48,900	60,500	9,100
2						
3	Sft^3/min (60°F)	300	8,800			
4	Temperature, $^\circ$F		1,250			
5	Pressure, psig					
6						
7						
8						
9						
10						

(Continued)

TABLE 19. (Continued)

	Stream No.	21	22	23	24	25
	Description	Settling tank overflow to surge tank	Settling tank solids return to dewatering bin	Reagent to surge tank	Makeup water	Water to bottom ash sluice
1	Total stream, lb/hr	42,000	18,500	200	700	40,700
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	26	27	28	29	30
	Description	Surge tank underflow to dewatering bins	Baghouse solids to disposal	Baghouse solids to transfer station	Solids to solids recycle silo	Feed slurry to spray dryer
1	Total stream, lb/hr	2,200	40,300	101,700	61,400	273,300
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	31	32	33	34	35
	Description	Makeup water to combined feed tank	Lime slurry to combined feed tank	Recycle slurry to combined feed tank	Makeup water to reslurry tank	Reslurried solids to recycle tank
1	Total stream, lb/hr	20,000	14,500	238,800	51,200	85,400
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	36	37	38		
	Description	Lime feed to wet ball mills	Makeup water to wet ball mills	Makeup water to recycle slurry tank		
1	Total stream, lb/hr	3,435	11,065	92,200		
2						
3	Sft^3/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

TABLE 20. CASE 2 EQUIPMENT LIST

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 1--Ammonia Storage and Injection		
1. Compressor, NH_3 unloading (2): 14.6 ft^3/min, capable of 250 psig suction max., 5-hp motor, cast iron body, insulated (1 operating, 1 spare)	8,600	2,100
2. Tank, NH_3 storage (5): Horizontal, 9-ft diameter x 66 ft long, 30,000 gal, 250 psig, carbon steel	169,500	3,900
3. Vaporizer, NH_3 (50): Electric resistance heaters, carbon steel shell, 15-kW rated, 10 per ammonia storage tank	29,800	800
4. Tank, ammonia accumulator (1): 281 ft^3, 5.5-ft diameter x 10 ft long, carbon steel, insulated (3 in.), +2.75-ft hemispherical end, 15 psig design pressure, 180°F design temperature	5,100	5,100
5. Ammonia absorber (1): 4 ft high x 1.1-ft diameter, 1.5-ft support, with vent, water supply, 1/4-in. Cor-Ten	400	1,900
6. Blower, air (3): 4,400 aft^3/min, 20 in. H_2O, 20 hp, carbon steel, insulated (2 operating, 1 spare)	12,500	1,400
7. Heater, dilution air (2): Fin tube steam heater, 500-ft^2 surface area, aluminum tubes, galvanized cabinet	26,200	800
8. Mixer, ammonia and dilution air (2): 30-in. diameter x 9.5 ft long, carbon steel	11,600	7,200
9. Injection grid, NH_3 and air (2): 32 ft wide, 19 ft high, Cor-Ten pipe and supports	90,700	96,600
10. Mixing grid, NH_3, air, and flue gas (2): 33 ft wide, 15 ft high, Cor-Ten pipe	14,000	27,600
Total, Area 1	368,400	147,400

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 2--Reactor		
1. Reactor (2): 53 ft wide x 37 ft long x 43 ft high, 6-in. mineral wool insulation; carbon steel housing, internals, and supports; elevated 40 ft	2,649,600	2,798,600
2. Sootblower, steam (20): 53 ft, retractable, 40-lb/min steam at 86 psig, 1 hp	686,400	33,100
3. Reactor crane and hoist (2): Electric 2-speed hoist, 2-ton capacity, 80-ft lift, grade to access door, 3 hp	21,200	600
4. Reactor hoist (4): Electric single-speed hoist, 2-ton capacity, access door to inside reactor, 3 hp	28,200	1,700
Total, Area 2	3,385,400	2,834,000
Area 3--Flue Gas-Handling Modifications[a]		
1. Fan, flue gas (2): Induced draft, 1,031,915 aft^3/min, ΔP = 20 in. H_2O, carbon steel, 5,000 hp, fluid drive, double width, double inlet	128,000	1,800
Total, Area 3	128,000	1,800
Area 4--Air Heater Modifications[b]		
1. Air heater (2): Vertical inverted, size 33, Ljungstrom type, Hot elements: DL type, 22 gauge, low alloy corrosion resistant, 26 in. deep, 206,700-ft^2 area	1,026,000	23,900

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Cold elements: NF type, 3.5-mm spacing, 22 gauge, low alloy corrosion resistant, 42 in. deep, 391,400-ft^2 area		
2. Sootblower, steam (2): Retractable, 175 lb/min steam at 200 psig	15,200	900
3. Pump, wash water booster (3): Centrifugal, 2,970 gpm, 210-ft head, 300 hp, carbon steel (2 operating, 1 spare)	3,900	900
Total, Area 4	1,045,100	25,700
Area 5--Waste Disposal[c]		
1. Landfill site development and construction (1): 75-acre landfill site, 1,475-ft square landfill, 3,213,000-yd^3 volume, 30-yr life, 98 ft high at center, 6,046-ft perimeter ditch to 61,000-yd^3 catchment basin	22,300	2,900
2. Wheel loader (1): 5.3-yd^3 bucket, diesel engine	2,200	
3. Dozer (1): Track type with straight blade, 103-hp diesel engine	1,000	
4. Compactor (1): Vibratory sheepsfoot compactor, self-propelled	1,300	
5. Wheel loader (1): 2.6-yd^3 bucket, diesel engine	1,000	
6. Water truck (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	300	

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
7. Service truck (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	400	
8. Onsite trailer for sanitary facilities and break room (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply, and 120-volt electric supply	100	
9. Onsite water supply and discharge treatment system (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps	300	200
10. Truck (2): Tandem-axle, 4-rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesek engine (1 operating, 1 spare), 0.7% of total truck costs in this area	900	
Total, Area 5	29,800	3,100

a. Costs shown are additional costs of boiler I.D. fan due to NO_x reactor pressure loss.

b. Costs shown are for modifications and additional equipment made necessary by NO_x removal.

c. Except as noted, 1.0% of total waste disposal costs is charged to NO_x removal.

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 6--Materials Handling		
1. Car shaker and crane (1): Top mounted with crane, 20-hp shaker, 7-1/2-hp hoist	71,900	13,000
2. Car puller (1): 25-hp puller, 5-hp return	63,000	19,600
3. Hopper, unloading (1): 16-ft diameter, 10-ft straight side, includes 6-in. square grating	15,500	5,900
4. Feeder, unloading (1): Vibrating, hopper mounted, 3-1/2-hp motor	3,800	300
5. Conveyor, lime unloading (1): Belt, 100 ton/hr, 20-ft horizontal, 5-hp motor	11,400	1,400
6. Dust collector, lime unloading pit (1): Bag filter, polypropylene bag, includes dust hoods, reverse jet cleaning	11,200	5,200
7. Elevator, lime storage silo (1): 100 ton/hr, 100 ft high, 50-hp motor	51,700	3,700
8. Concrete silo, lime storage (1): 44,960 ft3, 33.7-ft diameter, 50.5-ft straight side storage height, 30-day storage	82,500	172,100
9. Hopper bottom, lime storage silo (1): 60-degree cone, carbon steel	10,700	7,300
10. Feeder, lime storage silo reclaim (1): Hopper mounted, 3-1/2 hp, vibrating type	3,800	400
11. Conveyor, lime reclaim (1): Belt, 109-ft horizontal, 5-hp motor	23,900	3,100
12. Elevator, lime feed bin (1): 50 ft high, 50-hp motor	47,700	2,200

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
13. Bin, lime feed (2): 10-ft diameter, 15-ft straight side height, covered, carbon steel, vent filter	9,700	6,700
Total, Area 6	406,800	240,900
Area 7--Feed Preparation		
1. Feeder, lime bin unloading (2): Vibrating, 24 in. wide x 48 in. long, 16 ton/hr, 5 hp, carbon steel	8,300	600
2. Feeder, lime feed (2): Screw, 6-in. diameter x 12 ft long, 1 hp, 2 ton/hr	4,000	3,400
3. Slaker (2): Ball mill type, spiral classifier, mild steel, 2-ton/hr system, 4-ft inside diameter x 4-ft-long ball mill, 25 hp for mill, 14% manganese steel shell (1 operating, 1 spare)	107,100	14,000
4. Pump, lime slaker water supply (1): Centrifugal, 23 gpm, 100-ft head, 1 hp	1,000	600
5. Tank, slaker product (1): 6-ft diameter, 7 ft high, 1,450 gal, open top, four 6-in. baffles, agitator supports, carbon steel, neoprene lined	2,000	1,700
6. Agitator, slaker product tank (1): 2 turbines, 24-in. diameter, 2.5 hp, neoprene coated	8,000	900
7. Pump, slaker product tank (3): Centrifugal, 70 gpm, 150-ft head, 3 hp, carbon steel, neoprene lined (2 operating, 1 spare)	10,000	2,300
8. Tank, combined feed (1): 40-ft diameter x 20 ft high, 182,700 gal, open top, four 40-in. baffles, agitator supports, carbon steel, neoprene lined	46,300	37,600

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
9. Agitator, combined feed tank (1): 160-in. diameter, 50 hp, neoprene coated	60,900	4,600
10. Pump, combined feed tank (12): Centrifugal, 142 gpm, 100-ft head, 10 hp, carbon steel, neoprene lined (6 operating, 6 spares)	53,300	23,300
11. Pump, makeup water and dilution water for temperature control (1): Centrifugal, 60 gpm, 250-ft head, 15 hp, carbon steel	4,800	800
12. Pump, emergency flue gas quench (4): Centrifugal, 525 gpm, 200-ft head, 50 hp, carbon steel (3 operating, 1 spare)	28,500	3,800
13. Tank, overflow feed (4): 30 gal, 1.1-ft diameter x 4 ft high, neoprene lined	500	800
14. Dust collecting system (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, 3 hoods	7,800	2,700
Total, Area 7	342,500	97,100
Area 8--Flue Gas Handling[a]		
1. Fan, flue gas (4): Induced draft, 455,000 aft^3/min, ΔP = 12 in. H_2O, Inconel 625, 1,250 hp, fluid drive	1,199,100	19,700
Total, Area 8	1,199,100	19,700

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 9--SO_2 Absorption		
1. Spray dryer (4): 46-ft diameter x 41 ft high, straight side, carbon steel, 60-degree cone bottom, 40 ft long, 7-ft penthouse, total height 87 ft, one rotary atomizer per spray dryer, 700-hp motor on atomizer, (3 operating 1 spare)	7,846,300	1,011,200
Total, Area 9	7,846,300	1,011,200
Area 10--Lime Particulate Recycle		
1. Silo, solids recycle (2): 25-ft diameter x 37-ft straight side, 18,000 ft3, covered, carbon steel, porous stone air slide bin activator system	333,600	197,000
2. Vibrator (2): Bin actuator, 10-ft diameter, 5 hp	28,900	4,800
3. Pneumatic pressure transfer system (1): 12 in., 25 ton/hr, 1,000 ft long		
a. Conveying lines, pressure pneumatic for spray dryer solids (1): Pipelines and pipe fittings for pressure pneumatic conveyance of ash, 25-ton/hr conveying capacity with 1,000-ft equivalent length system, 10-in. I.D. branch lines and 12-in. I.D. main lines, nickel-chromium cast iron pipe with Ni-Hard or equivalent pipe fittings	25,300	13,600
b. Pressure feeders, ash and air (4): Materials-handling valve, electrically actuated, air operated, 10-in. I.D. ash inlet, 10-in. I.D. ash outlet, cast iron body, stainless steel slide gate; each assembly includes two spring-loaded, air-inlet check valves with cast iron bodies	32,000	18,200

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
c. Valves, line segregating (2): Segregating slide valve, electrically actuated, air operated for on-off control of each branch conveying line, 12-in. I.D. port, cast iron body, stainless steel slide gate	4,800	2,800
d. System control unit (1): Automatic sequence control unit to control the programmed operation of materials-handling valves, line segregating valves, and blowers; includes gauges for manual reading and override switches for manual operation	29,100	16,800
e. Filters, silo bag (2): Automatic cycling vent filter located on storage silo to remove residual ash from silo air discharge, 720-ft^2 bag area, 6 ft x 5.3 ft x 11 ft overall dimensions	27,700	15,900
f. Fans, bag filter vent (2): 2,364 aft3/min, 0.2 psig, 5-hp motor	12,700	7,200
4. Feeder, recycle slurry tank (2): Screw, 20-in. diameter x 17 ft long, 5 hp, 100 ton/hr	14,900	3,600
5. Tank, recycle slurry (1): 30-ft diameter x 31 ft high, 162,600 gal, open top, four 30-in. baffles, agitator support, carbon steel, neoprene lined	44,300	35,400
6. Agitator, recycle slurry tank (1): 120-in. diameter, 75 hp, neoprene coated	78,700	5,900
7. Pump, recycle slurry tank (3): Centrifugal, 308 gpm, 100-ft head, 25 hp, carbon steel, neoprene lined, 1 train of 3 pumps in series (2 operating, 1 spare)	15,200	6,200

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
8. Pump, makeup water (2): Centrifugal, 286 gpm, 150-ft head, 20 hp, carbon steel, (1 operating, 1 spare)	10,400	1,600
9. Conveyor, dragline (incline) (4): 50 ft/min, 24 in. wide x 15 ft long, 2 hp	120,000	3,300
10. Compressors, pneumatic pressure transfer system (1): 3,200 aft3/min, 15 psig, carbon steel, with silencers, 500-hp motor	34,700	3,100
11. Tank, reslurry (1): 15-ft diameter x 10 ft high, 11,900 gal, carbon steel, neoprene lined	7,300	5,800
12. Agitator, reslurry tank (1): 60-in. diameter, 15 hp, neoprene coated	25,600	1,900
13. Pump, reslurry tank (2): Centrifugal, 160 gpm, 100-ft head, 15 hp, carbon steel, neoprene lined	8,500	1,900
14. Conveyor, reslurry tank and reject stack feed (2): Belt, 285 ft long x 24 in. wide, 17 ton/hr, 2 hp	100,300	13,400
Total, Area 10	954,000	358,400
Area 11--Waste Disposal[b]		
1. Landfill site development and construction (1): 75-acre landfill site, 1,475-ft square landfill, 3,213,00-yd3 volume, 30-yr life, 98 ft high at center, 6,046-ft perimeter ditch to 61,000-yd3 catchment basin	345,200	44,500
2. Wheel loader (1): 5.3-yd3 bucket, diesel engine	33,700	

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
3. Dozers (1): Track type with straight blade, 103-hp diesel engine	15,000	
4. Compactor (1): Vibratory sheepsfoot compactor, self-propelled	20,700	
5. Wheel loader (1): 2.6-yd3 bucket, diesel engine	15,600	
6. Water truck (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	5,200	
7. Service truck (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	5,400	
8. Onsite trailer for sanitary facilities and break room (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply, and 120-volt electric supply	1,100	
9. Onsite water supply and discharge treatment system (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps	4,000	3,300

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
10. **Trucks** (2): Tandem-axle, 4-rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesel engine (1 operating, 1 spare), 15.7% of total truck costs in this area	20,900	
Total, Area 11	466,800	47,800

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a. These fans serve both SO_x and particulate removal cases. 33% of total fan costs is charged to SO_x removal for spray dryer pressure loss.

b. Except as noted, 15.9% of total waste disposal costs is charged to SO_2 removal.

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 12--Particulate Removal and Storage		
1. Baghouse, flue gas particulate removal (2): 880,200 aft3/min, automatic fabric filters, 1.76 gross air-to-cloth ratio, 500,115-ft^2 gross bag area, based on 80% availability, 2.2 net air-to-cloth ratio, 5-in. pressure drop, 28 compartments with 375 bags/compartment, 99.88% removal, 14 compartments/baghouse, each train 204 ft deep x 58 ft wide x 70 ft high (inside dimensions)	2,502,000	2,218,800
2. Hopper, economizer ash (10): Inverted pyramid-type double-V hopper, 15 ft x 7.7 ft wide x 7.4 ft deep, thermally isolated design, constructed of 3/8-in. Cor-Ten plate, 55-degree valley angle, each hopper has 2 outlets, 232-ft^3 volume and 252-ft^2 area per hopper	188,200	118,200
3. Hopper, air heater ash (10): Inverted pyramid-type double-V hopper, 15 ft long x 7.7 ft wide x 7.4 ft deep, constructed of 3/8-in. Cor-Ten plate, heat traced and insulated, 55-degree valley angle, each hopper has 2 outlets, 232-ft^3 volume and 252-ft^2 area per hopper, 6-kW heater	99,500	57,900
4. Hopper, baghouse particulates (56): Inverted pyramid-type double-V hopper, 29 ft long x 14.6 ft wide x 14 ft deep, constructed of 3/8-in. Cor-Ten plate, heat traced and insulated, each hopper has 2 outlets, 55-degree valley angle, 1,585-ft^3 volume and 917-ft^2 area per hopper, 10-kW heater	1,748,500	976,700
5. Hopper, NO$_x$ reactor ash (20): Inverted pyramid-type double-V hopper, 18.5 ft long x 10.6 ft wide x 10.2 ft deep, constructed of 3/8-in. Cor-Ten plate, insulated, 55-degree valley angle, each hopper has 2 outlets, 586-ft^3 volume and 448-ft^2 area per hopper, 10-kW heater	323,700	184,500

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
6. Hopper, bottom ash (1): 51 ft long x 10 ft wide x 9-1/2 ft high (inside dimensions), double-V hopper, center discharge with 2,170-ft3 capacity for 12-hr ash containment, supported independently of furnace-boiler, 3/8-in. carbon steel plate, refractory lined, 4 hydraulically operated exit doors emptying to 4 double-roll clinker grinders, 10-in. diameter x 2-ft-long manganese steel rolls, 60 hp	285,000	164,700
Total, Area 12	5,146,900	3,720,800
Area 13--Particulate Transfer		
1. Vacuum/pressure pneumatic fly ash and baghouse ash transfer system consisting of (1):		
a. Conveying lines, vacuum/pressure pneumatic for fly ashes and spray dryer solids (4): Pipelines and pipe fittings for vacuum pressure conveyance of ash from point of collection to transfer stations, 25-ton/hr conveying capacity with 1,320-ft equivalent length, 10-in. I.D. branch lines and 12-in. I.D. main lines, nickel-chromium cast iron pipe with Ni-Hard or equivalent pipe fittings	215,000	103,700
b. Valves, ash and air inlet (192): Materials-handling valve, electrically actuated, air operated, 10-in. I.D. ash inlet, 10-in. I.D. ash outlet cast iron body, stainless steel slide gate; each assembly includes two spring-loaded, air-inlet check valves with cast iron bodies	614,400	357,700
c. Valves, line segregating (24): Segregating slide valve, electrically actuated, air operated for on-off control of each branch conveying line, 12-in. I.D. port, cast iron body, stainless steel slide gate	86,400	6,900

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
d. Ash separation system comprising of (2): Primary air-ash separator (2): Primary centrifugal separator with tangential air-ash inlet, cyclone-type vortex finding sleeve, and top vertical outlet; two-gate, three-chamber ash removal and air-lock provision cycled for continuous vacuum operation; 5-ft diameter x 17 ft high; 40-ton/hr capacity, carbon steel shell, Ni-Hard in high-velocity compartment Secondary air-ash separator (2): Secondary centrifugal separator similar to primary unit except 3.5-ft diameter x 12 ft high for 6.9-ton/hr capacity Air-ash bag filter (2): Bag filter for air-ash service at 150°F, 19-in. Hg vacuum, 1,200-ft2 cloth area, cycled bag shaker and air-lock delivery to storage bin, 1.4-ton/hr capacity	96,000	66,600
e. Mechanical exhausters for supplying vacuum (6): Two-impeller, straight-lobe type, 2,000 aft3/min at 18-in. Hg vacuum and 150°F, 8-in. I.D. inlet connected to a common vacuum plenum, equipped with silencer, noise insulation, and inline prefilter, 200 hp (4 operating, 2 spares)	240,000	62,100
f. Transfer stations (4): Vacuum/pressure ash transfer units to convert from vacuum conveying medium to pressure conveying medium, 25 ton/hr/station, each station contains a primary and secondary cyclone, a filter, and the vacuum/pressure air-lock feeder, Ni-Hard or equivalent hardness wear surfaces, 5 hp (1-hr retention time)	320,000	182,100
g. Compressors (6): 2,000 aft3/min, 11 psig, 200 hp, intake filters, carbon steel body, silencers and noise insulation for body portion (4 operating, 2 spares)	240,000	18,600

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
h. Silo, solids storage (2): 30-ft diameter x 40 ft high, 28,300-ft^3 volume, with bin air fluidizing system, elevated construction for 11-1/2-ft truck clearance, rotary star feeders, carbon steel plate, 2 hp	455,000	257,700
i. Filter, silo bag (2): Automatic cycling vent filter, 2,436-ft^2 bag area, 20 ft x 5.3 ft x 11 ft overall dimensions	44,000	25,700
j. Fans, bag filter vent (2): 8,000 aft^3/min, ΔP = 6 in. H_2O, 20-hp motor	16,000	9,100
k. System control unit (1): Automatic sequence controller for the vacuum inlets and pressure stations, controls hopper levels, valve sequencing, alarms, blower operation, and other system monitoring; includes gauges for manual readings and override switches for manual operation	160,000	91,100
2. Bottom ash sluice transfer system (1):		
a. Pump, bottom ash water supply (3): Centrifugal, 255 gpm, 90-ft head, carbon steel, 10 hp (2 operating, 1 spare)	10,400	1,800
b. Pump, bottom ash water supply (3): Centrifugal, 1,860 gpm, 115-ft head, carbon steel, 75 hp (2 operating, 1 spare)	25,900	4,000
c. Pump, bottom ash water supply (3): Centrifugal, 587 gpm, 577-ft head, carbon steel, 150 hp (2 operating, 1 spare)	40,600	5,200

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
d. Tank, overflow (1): 18-ft diameter x 8 ft high, 11,400 gal, flat bottom, open top, with an overflow weir 2 ft below top of tank, 3/8-in. carbon steel, epoxide-coated interior	11,400	11,000
e. Pump, bottom ash hopper overflow bin (3): Centrifugal, 550 gpm, 175-ft head, carbon steel, 40 hp (2 operating, 1 spare)	17,400	2,600
f. Jet pump, bottom ash conveyance (4): Jet ejector nozzle assembly and adapter to bottom ash hopper, 360 gpm, 692-ft head supply water, Ni-Hard nozzle and throat construction (2 operating, 2 spares)	4,000	1,600
g. Sump pit, sluice (1): Concrete pit 5 ft wide x 5 ft long x 8 ft deep with two agitator nozzles located in bottom of bin to prevent settling	2,900	6,300
h. Pump, bottom ash sluice (3): Centrifugal slurry pumps, 2,550 gpm, 230-ft head, Ni-Hard liner and impeller, 250 hp (2 operating, 1 spare)	108,100	6,200
i. Valves, shutoff and crossover (17): Air-operated gate valve, 8-in. I.D. port, Ni-Hard	30,400	17,400
j. Slurry pipeline, one-quarter mile basalt-lined to dewatering bins, normal use (1): Pipeline comprising 74, 18-ft-long sections of flanged, basalt-lined steel pipe, 8-in. I.D. and 4 basalt-lined elbows or bends, 8-in. I.D.	82,000	29,500

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
k. Slurry pipeline, spare line to dewatering bins and return waterline (2): Pipeline comprising 34, 40-ft-long sections of flanged steel pipe, 8-in. I.D., schedule 80 carbon steel and 4 hardened elbows or bends, 8-in. I.D.	31,200	11,000
l. Bin, bottom ash dewatering (2): Conical-bottom dewatering bin, 25-ft diameter x 62 ft high, with 24-ft cylindrical section, 18-1/2-ft-high cone, 11,190-ft3 volume, stainless steel floating decanter and movable drainpipe, stationary decanter in conical section, erected for 16-1/2-ft truck clearance, carbon steel with stainless steel decanter drums, 250-ton capacity	200,000	93,600
m. Settling tank, bottom ash return water (1): 45-ft diameter x 13 ft deep, 154,800 gal, carbon steel, epoxide-coated interior, open top	68,000	38,900
n. Surge tank, bottom ash return water (1): Water reservoir, 35-ft diameter x 14 ft deep, 110,000 gal, carbon steel, epoxide-coated interior, open top	48,000	27,300
o. Pump, underflow solids recycle (3): Centrifugal, 250 gpm, 100-ft head, Ni-Hard steel body and impeller, 15 hp (2 operating, 1 spare)	11,300	2,300
p. Pump, dewatering bin sump pit (3): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined (2 operating, 1 spare)	7,200	2,500
3. Water treatment system for recycle water alkalinity control (1):		

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
a. Sulfuric acid storage tank for pH control of water (1): Cylindrical steel tank, 5-ft, 7-in. diameter x 5 ft, 7 in. high, 1,000 gal, flat bottom and closed flat top, carbon steel; all-weather housing	1,900	300
b. Metering pump for sulfuric acid (2): Positive displacement metering pump, 0.01 to 1 gpm, 0 psig, with flow rate controlled by a pH controller, Carpenter 20 alloy or similar corrosion resistance to 93% sulfuric acid; 0.25 hp (1 operating, 1 spare)	1,900	600
c. Agitator, treated water (1): Agitator with 24-in. diameter, nickel-chromium blade, 5 hp	2,900	400
Total, Area 13	3,192,300	1,443,800
Area 14--Flue Gas Handling[a]		
1. Fan, flue gas (4): Induced draft, 455,000 aft^3/min, ΔP = 12 in. H_2O, Inconel 625, 1,250-hp motor, fluid drive	2,398,200	39,300
Total, Area 14	2,398,200	39,300
Area 15--Waste Disposal[b]		
1. Landfill site development and construction (1): 75-acre landfill site, 1,475-ft square landfill, 3,213,000-yd^3 volume, 30-yr life, 98 ft high at center, 6,046-ft perimeter ditch to 61,000-yd^3 catchment basin	1,799,500	231,600

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
2. Wheel loader (1): 5.3-yd3 bucket, diesel engine	175,700	
3. Dozer (1): Track type with straight blade, 103-hp diesel engine	78,200	
4. Compactor (1): Vibratory sheepsfoot compactor, self-propelled	108,000	
5. Wheel loader (1): 2.6-yd3 bucket, diesel engine	81,500	
6. Water truck (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	27,000	
7. Service truck (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	28,200	
8. Onsite trailer for sanitary facilities and break room (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply, and 120-volt electric supply	5,600	
9. Onsite water supply and discharge treatment system (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps	21,000	17,200

(Continued)

TABLE 20. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
10. Truck (2): Tandem-axle, 4-rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesel engine (1 operating, 1 spare), 83.6% of total truck costs in this area	111,000	
Total, Area 15	2,435,700	248,800

a. These fans serve both SO_x and particulate removal areas. 67% of total fan costs is charged to particulate removal for baghouse pressure loss.

b. Except as noted, 83.1% of total waste disposal costs is charged to ash disposal.

TABLE 21. STEAM SOOTBLOWING AND WATER WASHING REQUIREMENTS

FOR AIR HEATERS OF CASE 2

	Steam sootblowing					
Case	Number of blowers	Cycles/day/ blower	Min/cycle	Lb steam/min	Lb steam/yr	Additional lb steam/yr
Standard without SCR	2	3	20	127	3,492,500	
Modified with SCR	4	3	22	175	10,587,500	7,095,000

	Water washing					
Case	Cycles/yr	Hr/cycle	Gal/ min/heater	Psig	Gal/yr	Additional gal/yr
Standard without SCR	1	8	2,440	150	2,342,400	
Modified with SCR	2	8	2,970	150	5,702,400	3,360,000

Area 5 - Waste Disposal--

The waste disposal area description is identical to that of case 1. The spent catalyst volume is slightly larger because of the larger volume of catalyst used in case 2.

SO_2 Control

Processing areas 6 through 11 describe the spray dryer FGD process. Notice that there is no processing area for particulate collection in the SO_2 control process. Differentiation of costs and functions of fabric filter particulate control between SO_2 particulate collection and fly ash collection is impractical; therefore, all baghouse costs are included in the particulate control section.

Area 6 - Materials Handling--

The materials-handling area consists of the equipment to unload pebble quicklime from trucks or railcars, transfer it to concrete storage silos at the FGD system, and transfer it from the silos to the feed preparation area. The unloading and conveying system is designed for 100 ton/hr and the silos have a 30-day capacity. The lime is transferred by closed conveyor and loaded into the silos and feed bin by bucket elevator.

Area 7 - Feed Preparation--

The feed preparation area consists of equipment to slake the lime and prepare and meter the absorbent slurry to the spray dryers, along with other auxiliary equipment. The lime is metered to two parallel trains of ball mill slakers equipped with spiral classifiers and oversized particle recycle. The slaked lime slurry from both slakers is combined in an agitated slaker receiver tank as a 20% solids slurry. This slurry is combined with the 40% solids slurry from the lime particulate recycle area (area 10) in an 8-hour capacity combined feed tank, from which the combined slurry is metered to the spray dryer atomizers. The feed preparation area also meters dilution water to the atomizers for feed concentration control and emergency quench pumps to protect the baghouse in case of interruptions to the feed slurry addition.

Area 8 - Flue Gas Handling--

The flue gas-handling area contains the same general equipment and has the same function as the flue gas handling area of case 1. It consists of the incremental increase in the size of the boiler ID fan to compensate for the 4-inch H_2O pressure drop in the FGD system (the remaining incremental increase is prorated to processing area 14) and the FGD system ductwork. The ductwork consists of the spray dryer inlet plenum, the ducts connecting the spray dryers to the inlet plenum and to the baghouse plenum, the individual spray dryer bypass ducts, and the two emergency bypass ducts with a combined capacity of 50% of the scrubbed flue gas that connect the inlet plenum with the stack plenum.

Area 9 - SO_2 Absorption--

This area consists of the four spray dryers and their atomizer systems. Each spray dryer is a 46-foot-diameter carbon steel vessel with a total height of 87 feet, including a 60-degree conical bottom section. A single 700-hp rotary atomizer is mounted at the top of each spray dryer. The flue gas enters through a gas distribution manifold around the atomizer that imparts a swirling moment to the gas. The flue gas leaves the spray dryer through a downward-opening horizontal duct in the conical bottom. Larger particles collect in the conical bottom and are removed through an air lock. The spray dryers are designed for an overall excess SO_2 removal capacity of 33% so that three can meet the SO_2 removal requirements; normally, however, all four are operated to reduce the overall pressure drop.

The spray dryers are designed for an SO_2 removal efficiency of 73% at a stoichiometry of 0.79 mol $Ca(OH)_2$/mol SO_2 entering, a 10-second residence time, and an 18°F approach to saturation. At an entering flue gas temperature of 300°F, this produces an outlet temperature of 154°F. A bypass for 12% of the flue gas is provided for each spray dryer to ensure dry operation of the baghouse. The gas entering the baghouse, which consists of the treated and bypassed gas, has a temperature of 170°F.

Area 10 - Lime Particulate Recycle--

This area consists of equipment to convey, store, and reslurry particulate matter from the bottom of the spray dryers and some of the particulate matter collected in the baghouse. Each spray dryer hopper has an air lock and a drag chain conveyor that carries the solids to a common belt conveyor. The conveyor carries the solids to the reslurry tank at 34,100 lb/hr. A portion of the solids from the baghouse is pneumatic-pressure conveyed to the solids recycle silos and metered at a rate of 61,400 lb/hr to the recycle slurry tank. The total recycle rate is 95,500 lb/hr of solids as a 40% solids slurry. This corresponds to a 1 to 27.8 ratio of fresh lime to recycle solids.

Area 11 - Waste Disposal--

The FGD waste, collected commingled with the fly ash in the baghouse, is transported and disposed of in the common landfill as described in the particulate control section. Costs are prorated from the total transportation and disposal costs on the basis of volume.

Particulate Control

Processing areas 12 through 15 describe the bottom ash, fly ash, and fly ash - FGD solids control processes.

Area 12 - Particulate Removal and Storage--

The baghouse; all bottom ash, fly ash, and baghouse hoppers; the bottom ash transport and dewatering system; the pneumatic particulate transfer system; and the particulate storage silos are included in this area. Two baghouses in parallel are used, each 204 feet long, 58 feet wide, and 70 feet high (inside dimensions). Each baghouse is designed for a flow rate of 440,100 aft^3/min and has a gross air-to-cloth ratio of 1.76 $aft^3/min/ft^2$ and a net air-to-cloth ratio of 2.2 at a pressure drop of 5 inches H_2O. Each baghouse has 14 compartments with 375 bags per compartment. The design removal efficiency is 99.88%.

There are 96 individual double-vee hoppers in which dry particulate matter is collected: 56 on the baghouse, 10 on the economizer, 10 on the air heaters, and 20 on the SCR reactors. The description is the same as the description in case 1 (area 13).

Area 13 - Particulate Transfer--

The vacuum-pressure pneumatic system to remove the fly ash and fly ash - FGD solids from the hoppers and transfer it to storage silos and the bottom ash transfer and dewatering system are included in this area. The bottom ash system is identical in function to that of case 1 (area 15), differing only in size because of the smaller quantity of ash. The fly ash and fly ash - FGD particulates are transported with a combined vacuum-pressure pneumatic system because of the large number of hoppers involved and the resulting longer transport distances. The particulates are collected from the 192 hopper outlets with a vacuum system and transported to 4 transfer stations. From the transfer stations, they are transported by a pressure pneumatic system to two elevated carbon steel storage silos 30 feet in diameter and 40 feet high.

Area 14 - Flue Gas Handling--

This area consists of the plenum that distributes flue gas to the baghouses, the four ducts connecting the baghouses to the stack, and four ID booster fans to compensate for the 12-inch H_2O pressure drop through the FGD system and the baghouses (costs are prorated between the two systems based on an 8-inch H_2O pressure drop for the baghouses and their associated ductwork).

Area 15 - Waste Disposal--

Trucking of the combined fly ash - FGD waste and bottom ash to the landfill and landfill operations are included in this area. The costs are prorated based on the volume of spent SCR process catalyst, FGD waste, and ash.

CASE 3

Case 3 is based on 0.7% sulfur western subbituminous coal. It consists of a hot-side ESP, an SCR NO_x control process, and a limestone FGD process in which natural oxidation produces a 95% gypsum waste. The flow diagram is shown in Figure 5, the material balance is shown in Table 22, and the equipment list is shown in Table 23. Each process area is described below with the exception of those in which the verbal description is similar to those for the same areas previously described in cases 1 and 2.

NO_x Control

The 2-train SCR reactor system is essentially the same as the system in case 2 except for the differences that result from upstream fly ash removal: a different equipment layout to accommodate the hot-side ESP, a smaller catalyst volume because of reduced fly ash fouling, and elimination of fly ash hoppers on the reactors. Areas 1 through 5 describe the NO_x control process.

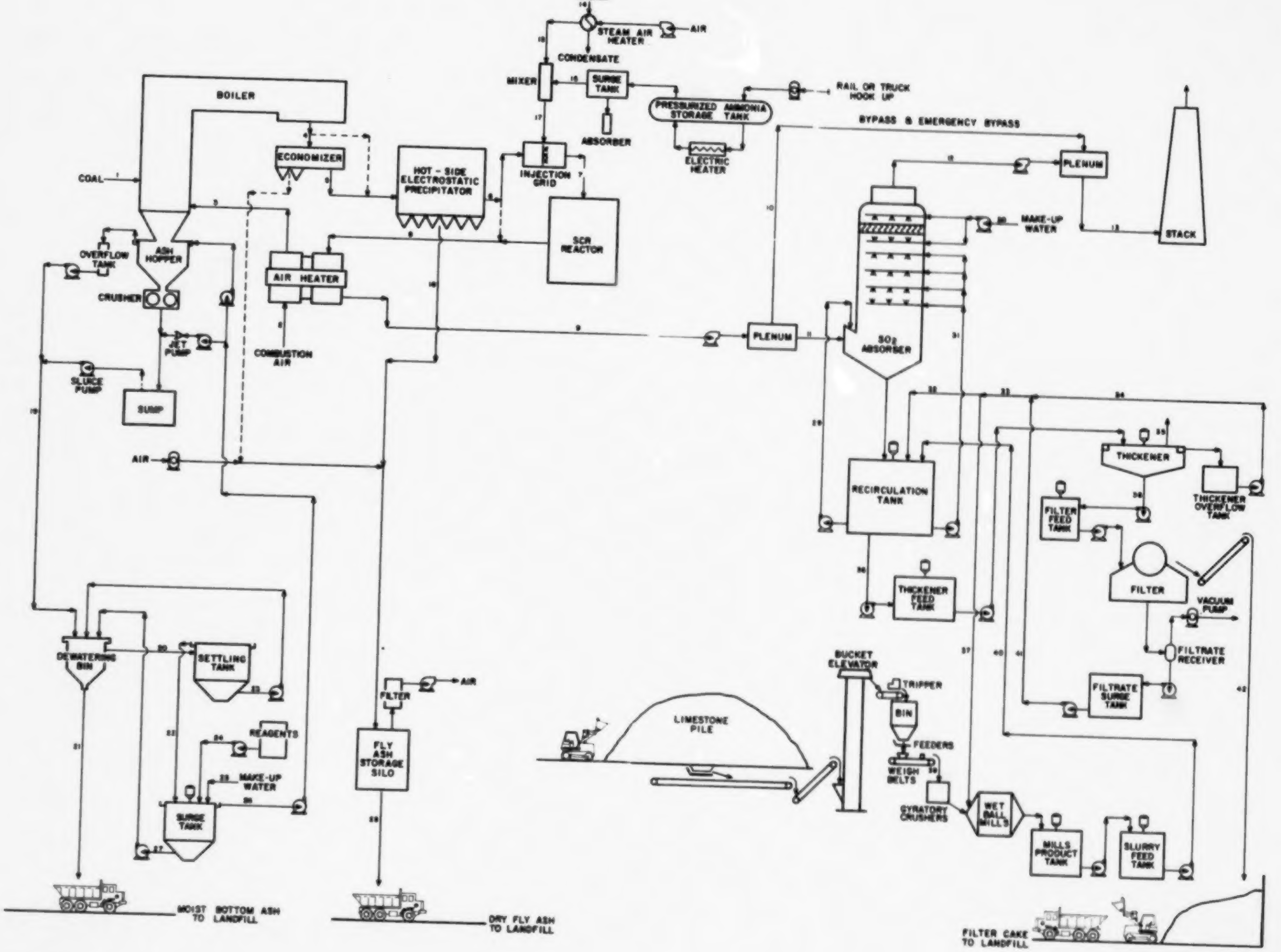
STEAM
STEAM AIR HEATER
AIR
CONDENSATE
MIXER
SURGE TANK
ABSORBER
RAIL OR TRUCK HOOK UP
PRESSURIZED AMMONIA STORAGE TANK
ELECTRIC HEATER
BYPASS & EMERGENCY BYPASS
BOILER
COAL
ECONOMIZER
HOT - SIDE ELECTROSTATIC PRECIPITATOR
INJECTION GRID
SCR REACTOR
PLENUM
STACK
MAKE-UP WATER
OVERFLOW TANK
ASH HOPPER
CRUSHER
AIR HEATER
COMBUSTION AIR
JET PUMP
SLUICE PUMP
SUMP
SO_2 ABSORBER
RECIRCULATION TANK
THICKENER
THICKENER OVERFLOW TANK
FILTER FEED TANK
FILTER
VACUUM PUMP
FILTRATE RECEIVER
FILTRATE SURGE TANK
THICKENER FEED TANK
DEWATERING BIN
SETTLING TANK
REAGENTS
MAKE-UP WATER
SURGE TANK
FILTER
FLY ASH STORAGE SILO
LIMESTONE PILE
BUCKET ELEVATOR
TRIPPER
BIN
FEEDERS
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
MILLS PRODUCT TANK
SLURRY FEED TANK
MOIST BOTTOM ASH TO LANDFILL
DRY FLY ASH TO LANDFILL
FILTER CAKE TO LANDFILL

Figure 5. Case 3 flow diagram.

TABLE 22. CASE 3 MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Combustion air to air heater	Combustion air to boiler	Gas to economizer	Gas to ESP
1	Total stream, lb/hr	640,200	5,765,200	4,977,200	5,609,200	5,609,200
2						
3	Sft^3/min (60°F)		1,273,900	1,099,800	1,215,300	1,215,300
4	Temperature, °F		80			750
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	6	7	8	9	10
	Description	Gas to ammonia injection grid	Gas with ammonia to SCR reactor	Gas to air heater	Gas to inlet plenum	Spray tower bypass gas
1	Total stream, lb/hr	5,576,800	5,616,000	5,616,000	6,404,000	1,800,200
2						
3	Sft^3/min (60°F)	1,215,300	1,224,100	1,224,200	1,398,300	393,100
4	Temperature, °F	750	747	750	300	300
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	11	12	13	14	15
	Description	Gas to spray tower	Gas from spray tower	Gas to stack	Steam to dilution air heater	Dilution air to mixer
1	Total stream, lb/hr	4,603,800	4,798,000	6,598,200	2,000	38,412
2						
3	Sft^3/min (60°F)	1,005,200	1,074,000	1,467,100		8,500
4	Temperature, °F	300	135		298	250
5	Pressure, psig				50	
6						
7						
8						
9						
10						

	Stream No.	16	17	18	19	20
	Description	Ammonia to mixer	Ammonia-air mixture to injection grid	Fly ash to storage silo	Bottom ash from boiler	Bottom ash sluice water to settling tank
1	Total stream, lb/hr	788	39,200	32,400	48,900	60,500
2						
3	Sft^3/min (60°F)	300	8,800			
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

(Continued)

TABLE 22. (Continued)

	Stream No.	21	22	23	24	25
	Description	Dewatered bottom ash to landfill	Settling tank overflow to surge tank	Settling tank solids return to dewatering bin	Reagent to surge tank	Makeup water
1	Total stream, lb/hr	9,100	42,000	18,500	200	700
2						
3	Sft^3/min (60°F)					
4	Temperature, $^\circ$F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	26	27	28	29	30
	Description	Water to bottom ash sluice	Surge tank underflow to dewatering bins	Fly ash to landfill	Recycle slurry to presaturator	Makeup water to spray tower
1	Total stream, lb/hr	40,700	2,200	32,400	2,580,000	205,600
2						
3	Sft^3/min (60°F)					
4	Temperature, $^\circ$F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	31	32	33	34	35
	Description	Recycle slurry to spray tower	Clear liquid to recirculation tank	Clear liquid return	Thickener overflow return	Thickener evaporation
1	Total stream, lb/hr	74,288,800	121,000	126,300	110,800	5,600
2						
3	Sft^3/min (60°F)					
4	Temperature					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.	36	37	38	39	40
	Description	Slurry to thickener	Clear liquid to wet ball mills	Thickener bottoms to filter	Limestone to wet ball mills	Limestone slurry to recirculation tank
1	Total stream, lb/hr	145,700	5,300	29,300	8,000	13,300
2						
3	Sft^3/min (60°F)					
4	Temperature, $^\circ$F					
5	Pressure, psig					
6						
7						
8						
9						
10						

(Continued)

TABLE 22. (Continued)

	Stream No.	41	42			
	Description	Filtrate return	Filter cake to landfill			
1	Total stream, lb/hr	15,500	13,800			
2						
3	Sft³/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.					
	Description					
1	Total stream, lb/hr					
2						
3	Sft³/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.					
	Description					
1	Total stream, lb/hr					
2						
3	Sft³/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

	Stream No.					
	Description					
1	Total stream, lb/hr					
2						
3	Sft³/min (60°F)					
4	Temperature, °F					
5	Pressure, psig					
6						
7						
8						
9						
10						

TABLE 23. CASE 3 EQUIPMENT LIST

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 1--Ammonia Storage and Injection		
1. Compressor, NH_3 unloading (2): 14.6 ft^3/min, capable of 250 psig suction max., 5-hp motor, cast iron body, insulated (1 operating, 1 spare)	8,600	2,100
2. Tank, NH_3 storage (5): Horizontal, 9-ft diameter x 66 ft long, 30,000 gal, 250 psig, carbon steel	169,500	3,900
3. Vaporizer, NH_3 (50): Carbon steel, 15 kW, electric resistance heater, 10 per ammonia storage tank	29,800	800
4. Tank, ammonia accumulator (1): 281 ft^3, 5.5-ft diameter x 10 ft long, carbon steel, insulated (3 in.), +2.75-ft hemispherical end, 15 psig design pressure, 180°F design temperature	5,100	5,100
5. Ammonia absorber (1): 4 ft high x 1.1-ft diameter, 1.5-ft support, with vent, water supply, 1/4-in. Cor-Ten	400	1,900
6. Blower, air (3): 4,400 aft^3/min, 20 in. H_2O, 20 hp, carbon steel, insulated (2 operating, 1 spare)	12,500	1,400
7. Heater, dilution air (2): Fin tube steam heater, 500-ft^2 surface area, aluminum tubes, galvanized cabinet	26,200	800
8. Mixer, ammonia and dilution air (2): 30-in. diameter x 9.5 ft long, carbon steel	11,600	7,200
9. Injection grid, NH_3 and air (2): 21 ft wide, 26 ft high, Cor-Ten pipe and supports	63,400	82,600
10. Mixing grid, NH_3, air, and flue gas (2): 22 ft wide, 22 ft high, Cor-Ten pipe	13,800	27,000
Total, Area 1	340,900	132,800

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 2--Reactor		
1. Reactor (2): 49 ft wide x 40 ft long x 42 ft high, 6-in. mineral wool insulation; carbon steel housing, internals, and supports; elevated 20 ft	2,437,200	2,580,900
2. Sootblower, steam (20): 49 ft, retractable, 35-lb/min steam at 86 psig, 1 hp	540,000	33,100
3. Reactor crane and hoist (2): Electric 2-speed hoist, 2-ton capacity, 60-ft lift, grade to access door, 3 hp	17,800	600
4. Reactor hoist (4): Electric single-speed hoist, 2-ton capacity, access door to inside reactor, 3 hp	28,200	1,700
Total, Area 2	3,023,200	2,616,300
Area 3--Flue Gas-Handling Modifications[a]		
1. Fan, flue gas (2): Induced draft, 1,031,915 aft^3/min, $\Delta P = 22$ in. H_2O, carbon steel, 5,000 hp, fluid drive, double width, double inlet	107,300	1,300
Total, Area 3	107,300	1,300
Area 4--Air Heater Modifications[b]		
1. Air heater (2): Vertical inverted, size 32.5, Ljungstrom type,	747,000	9,600
Hot elements: DN type, 22 gauge, low alloy corrosion resistant, 26 in. deep, 186,100-ft^2 area		

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Cold elements: NF type, 3.5-mm spacing, 22 gauge, low alloy corrosion resistant, 42 in. deep, 359,500-ft^2 area		
2. Sootblower, steam (2): Retractable, 127-lb/min steam at 200 psig	15,200	900
Total, Area 4	762,200	10,500
Area 5--Waste Disposal^c		
1. Landfill site development and construction (1): 80-acre landfill site, 1,533-ft square landfill, 3,559,000-yd^3 volume, 30-yr life, 101 ft high at center, 6,277-ft perimeter ditch to 66,000-yd^3 catchment basin	20,000	2,500
2. Wheel loader (1): 5.3-yd^3 bucket, diesel engine	1,800	
3. Dozer (1): Track type with straight blade, 109-hp diesel engine	900	
4. Compactor (1): Vibratory sheepsfoot compactor, self-propelled	1,200	
5. Wheel loader (1): 2.6-yd^3 bucket, diesel engine	800	
6. Water truck (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	300	
7. Service truck (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	300	

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
8. Onsite trailer for sanitary facilities and break room (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply, and 120-volt electric supply	100	
9. Onsite water supply and discharge treatment system (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps		
10. Truck (2): Tandem-axle, 4 rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesel engine (1 operating, 1 spare), 0.6% of total truck costs in this area	700	
Total, Area 5	26,300	2,700

a. Costs shown are additional costs of boiler I.D. fan due to NO_x reactor pressure loss.

b. Costs shown are for modifications and additional equipment necessary for NO_x removal.

c. Except as noted, 0.9% of total waste disposal costs is charged to NO_x removal.

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 6--Materials Handling		
1. Mobile equipment (1): Bucket tractor, 2-1/2-yd3 bucket, storage pile is 52,300 ft3	75,900	0
2. Hopper, reclaim (1): 7-ft diameter x 4-1/4 ft deep x 2-ft bottom, 75 ft3, carbon steel, 60-degree cone bottom	1,200	800
3. Feeder, reclaim (1): Vibrating pan, 3-1/2 hp, 100 ton/hr	5,500	500
4. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7.5 hp, reverse jet cleaning, includes dust hoods	7,800	2,600
5. Pump, reclaim sump pit (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400	800
6. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 100 ft long, 2 hp, 100 ton/hr, 105 ft/min	22,900	1,400
7. Conveyor, limestone reclaim (incline) (1): Belt, 30 in wide x 193 ft long, 40 hp, 15-degree incline, 50-ft lift, 100 ton/hr, 105 ft/min	60,300	3,700
8. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90-ft lift, 100 ton/hr	57,800	6,700
9. Conveyor, feed (1): Belt, 30 in. wide x 60 ft long, 7.5 hp, 100 ton/hr, 105 ft/min	20,500	1,400
10. Tripper, feed conveyor (1): 30 ft/min, 1 hp	27,200	9,100

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
11. Bin, crusher feed (3): 13-ft diameter x 21-ft straight side height, 3,100 ft3, covered, 50-degree cone, carbon steel	43,300	24,100
Total, Area 6	324,800	51,100
Area 7--Feed Preparation		
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp	49,600	2,300
2. Crusher (3): Gyratory, 75 hp	297,100	6,500
3. Ball mill, wet (3): Wet, open system, 7-ft diameter x 10-1/2 ft long, 113 hp, 2.0 ton/hr (2 operating, 1 spare)	503,200	61,600
4. Dust collector, ball mill (3): Bag filter, polypropylene bag, 2,200 aft3/min, 7.5 hp, reverse jet cleaning, 2 hoods	23,300	7,800
5. Tank, mills product (3): 10-ft diameter x 10 ft high, 5,500 gal, open top, four 10-in. baffles, agitator supports, carbon steel, glass-filled polyester lining	13,700	11,000
6. Agitator, mills product tank (3): 40-in. diameter, 10 hp, neoprene coated	22,900	5,500
7. Pump, mills product tank (3): Centrifugal 8 gpm, 60-ft head, 1 hp, carbon steel, neoprene lined (2 operating, 1 spare)	7,600	2,700
8. Tank, slurry feed (1): 11.5-ft diameter x 11.5 ft high, 8,800 gal, open top, four 12-in. baffles, agitator supports, carbon steel, glass-filled polyester lining	5,700	4,700

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
9. Agitator, slurry feed (1): 44-in. diameter, 14 hp, neoprene coated	11,400	900
10. Pump, slurry feed (6): Centrifugal, 6 gpm, 60-ft head, 1/4 hp, carbon steel, neoprene lined (3 operating, 3 spares)	14,900	5,500
Total, Area 7	949,400	108,500
Area 8--Flue Gas Handling		
1. Fan, flue gas (4): Induced draft, 409,800 aft3/min, ΔP = 7.2 in. H_2O, 663 hp, fluid drive, double width, double inlet, Inconel 625	2,775,400	48,900
Total, Area 8	2,775,400	48,900
Area 9--SO_2 Absorption		
1. SO_2 absorber (4): Spray tower, 40 ft x 37 ft wide x 18-1/2 ft deep, 1/4-in. carbon steel, neoprene lining, 316 stainless steel grids, FRP chevron vane entrainment separator, slurry header and nozzles	5,087,500	402,300
2. Tank, recirculation (4): 41.3-ft diameter x 41.3 ft high, 413,100 gal, open top, four 41-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining	337,600	272,800
3. Agitator, recirculation tank (4): 156-in. diameter, 78 hp, neoprene coated	421,300	138,200

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
4. Pump, presaturator (8): Centrifugal, 1,636 gpm, 100-ft head, 72 hp, carbon steel, neoprene lined (3 operating, 5 spares)	87,200	27,700
5. Pump, recirculation (16): Centrifugal, 15,600 gpm, 100-ft head, 692 hp, carbon steel, neoprene lined (9 operating, 7 spares)	1,647,200	148,800
6. Pump, makeup water (2): Centrifugal, 3,068 gpm, 200-ft head, 258 hp, carbon steel (1 operating, 1 spare)	30,400	3,400
7. Sootblower (32): Air, fixed	89,500	83,400
Total, Area 9	7,700,700	1,076,600
Area 10--Solids Separation		
1. Tank, thickener feed (1): 20.6-ft diameter x 41.3 ft high, 103,200 gal, open top, four 20-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining	33,600	27,800
2. Agitator, thickener feed tank (1): 78-in. diameter, 49 hp, neoprene coated	37,300	3,100
3. Pump, thickener feed (2): Centrifugal, 276 gpm, 60-ft head, 7 hp, carbon steel, neoprene lined (1 operating, 1 spare)	8,700	3,200
4. Thickener (1): 19-ft diameter x 4.6 ft high, carbon steel sides, concrete basin, includes 1/4-hp rake motor and mechanism, 272-ft^2 area	33,600	23,600
5. Tank, thickener overflow (1): 11.7-ft diameter x 4.6 ft high, 3,655 gal, open top, carbon steel	1,700	1,200

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
6. Pump, thickener overflow tank (2): Centrifugal, 222 gpm, 75-ft head, 7 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,300	1,000
7. Pump, thickener underflow (2): Centrifugal, 44 gpm, 5-ft head, 1/4 hp, carbon steel, neoprene lined (1 operating, 1 spare)	4,800	1,800
8. Tank, filter feed (1): 5-ft diameter x 5 ft high, 723 gal, open top, four 5-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining	1,100	900
9. Agitator, filter feed tank (1): 19-in. diameter, 2 hp, neoprene coated	900	100
10. Pump, filter feed tank (3): Centrifugal, 22 gpm, 50-ft head, 1 hp, carbon steel, neoprene lined (2 operating, 1 spare)	7,800	2,700
11. Filter (3): Rotary vacuum, 3-ft diameter x 6-ft face, 7 hp, includes auxiliary equipment, 58-ft^2 filtration area (2 operating, 1 spare)	163,300	47,600
12. Pump, filtrate (4): Centrifugal, 15 gpm, 20-ft head, 1/4 hp, carbon steel, neoprene lined (2 operating, 2 spares)	16,500	1,900
13. Tank, filtrate surge (1): 4.4-ft diameter x 4.4 ft high, 508 gal, open top, carbon steel	500	300
14. Pump, filtrate surge tank (2): Centrifugal, 31 gpm, 85-ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	8,400	1,000
15. Conveyor, filtrate cake (1): Belt, 30 in. wide x 75-ft-long horizontal, 1-1/2 hp, 7-1/2 ton/hr, 100-ft incline	37,100	3,500
Total, Area 10	364,600	119,700

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 11--Waste Disposal[a]		
1. Landfill site development and construction (1): 80-acre landfill site, 1,533-ft square landfill, 3,559,000-yd3 volume, 30-yr life, 101 ft high at center, 6,277-ft perimeter ditch to 66,000-yd3 catchment basin	562,500	69,400
2. Wheel loader (1): 5.3-yd3 bucket, diesel engine	51,200	
3. Dozer (1): Track type with straight blade, 109-hp diesel engine	24,200	
4. Compactor (1): Vibratory sheepsfoot compactor, self-propelled	33,700	
5. Wheel loader (1): 2.6-yd3 bucket, diesel engine	23,700	
6. Water truck (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	7,900	
7. Service truck (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	8,900	
8. Onsite trailer for sanitary facilities and break room (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply; and 12-volt electric supply	1,600	

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
19. Onsite water supply and discharge treatment system (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps	6,500	5,300
10. Truck (2): Tandem-axle, 4-rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesel engine (1 operating, 1 spare), 24.1% of total truck costs in this area	32,000	
Total, Area 11	752,200	74,700

a. Except as noted, 24.2% of total waste disposal costs is charged to SO_2 removal.

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 12--Particulate Removal and Storage		
1. Electrostatic precipitator, hot side (2): 1,447,150 aft3/min, 651,218-ft^2 collection area, 1-in. pressure drop, 99.52% removal efficiency, 450-ft^2/kaft3/min SCA, 57.8 ft deep x 85.7 ft wide x 46.3 ft high (inside dimensions)	4,694,500	3,856,500
2. Hopper, economizer ash (10): Inverted pyramid-type double-V hopper, 15 ft long x 7.7 ft wide x 7.4 ft deep, thermally isolated design, constructed of 3/8-in. Cor-Ten plate, 55-degree valley angle, each hopper has 2 outlets, 232-ft3 volume and 252-ft^2 area per hopper	188,200	118,200
3. Hopper, ESP ash (24): Inverted pyramid-type double-V hopper, 28.6 ft long x 14.5 ft wide x 13.9 ft deep, constructed of 3/8-in. Cor-Ten plate, heat traced and insulated, each hopper has 2 outlets, 55-degree valley angle, 1,545-ft3 volume and 899-ft^2 area per hopper, 10-kW heater	738,100	412,300
4. Hopper, bottom ash (1): 51 ft long x 10 ft wide x 9-1/2 ft high (inside dimensions), double-V hopper, center discharge with 2,170-ft3 capacity for 12-hr ash containment, supported independently of furnace-boiler, 3/8-in. carbon steel plate, refractory lined, 4 hydraulically operated exit doors emptying to 4 double-roll clinker grinders, 10-in. diameter x 2-ft-long manganese steel rolls, 60 hp	285,000	164,700
Total, Area 12	5,905,800	4,551,700
Area 13--Particulate Transfer		
1. Pressure pneumatic transfer system for fly ash (1):		

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
a. Conveying line, pressure pneumatic for fly ashes (1): Pipelines and pipe fittings for pressure pneumatic conveyance of ash, 35-ton/hr conveying capacity with 1,320-ft equivalent length system, 8-in. I.D. branch lines and 10-in. I.D. main lines, nickel-chromium cast iron pipe with Ni-Hard or equivalent pipe fittings	53,200	28,800
b. Pressure feeder, ash and air (68): Materials-handling valve, electrically actuated, air operated, 10-in. I.D. ash inlet, 10-in. I.D. ash outlet, cast iron body, stainless steel slide gate; each assembly includes two spring-loaded, air-inlet check valves with cast iron bodies	544,000	309,700
c. Valve, line segregating (9): Segregating slide valve, electrically actuated, air operated for on-off control of each branch conveying line, 10-in. I.D. port, cast iron body, stainless steel slide gate	20,000	11,600
d. System control unit (1): Automatic sequence control unit to control the programmed operation of materials-handling valves, line segregating valves, and blowers; includes gauges for manual reading and override switches for manual operation	80,000	45,500
e. Filter, silo bag (2): Automatic cycling vent filter, 1,120-ft^2 bag area, 9.3 ft x 5.3 ft x 11 ft overall dimensions	36,000	20,600
f. Fan, bag filter vent (2): 3,224 aft^3/min, .20 psig, 5 hp	12,800	7,500
g. Compressor, pressure pneumatic transfer system (3): 3,200 aft^3/min, 15 psig, 500-hp motor, carbon steel, with silencers (2 operating, 1 spare)	104,000	9,300

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
h. Silo, fly ash storage (2): 30-ft diameter x 32 ft high, 22,600-ft3 volume, with bin air fluidizing system, elevated construction for 11-1/2-ft truck clearance, rotary star feeders, carbon steel plate, 2 hp	375,000	212,200
2. Bottom ash sluice transfer system (1):		
a. Pump, bottom ash water supply (3): Centrifugal, 255 gpm, 90-ft head, carbon steel, 10 hp (2 operating, 1 spare)	10,400	1,800
b. Pump, bottom ash water supply (3): Centrifugal, 1,860 gpm, 115-ft head, carbon steel, 75 hp (2 operating, 1 spare)	25,900	4,000
c. Pump, bottom ash water supply (3): Centrifugal, 587 gpm, 577-ft head, carbon steel, 150 hp (2 operating, 1 spare)	40,600	5,200
d. Tank, overflow (1): 18-ft diameter x 8 ft high, 11,400 gal flat bottom, open top, with an overflow weir 2 ft below top of tank, 3/8-in. carbon steel, epoxide-coated interior	11,400	11,000
e. Pump, bottom ash hopper overflow bin (3): Centrifugal, 550 gpm, 175-ft head, carbon steel, 40 hp (2 operating, 1 spare)	17,400	2,600
f. Jet pump, bottom ash conveyance (4): Jet ejector nozzle assembly and adapter to bottom ash hopper, 360 gpm, 692-ft head supply water, Ni-Hard nozzle and throat construction (2 operating, 2 spares)	4,000	1,600
g. Sump pit, sluice (1): Concrete pit, 5 ft wide x 5 ft long x 8 ft deep with two agitator nozzles located in bottom of bin to prevent settling	2,900	6,300

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
h. Pump, bottom ash sluice (3): Centrifugal slurry pumps, 2,550 gpm, 230-ft head, Ni-Hard liner and impeller, 250 hp (2 operating, 1 spare)	108,100	6,200
i. Valve, shutoff and crossover (17): Air-operated gate valve, 8-in. I.D. port, Ni-Hard	30,400	17,400
j. Slurry pipeline, one-quarter mile basalt-lined to dewatering bins, normal use (1): Pipeline comprising 74, 18-ft-long sections of flanged, basalt-lined steel pipe, 8-in. I.D. and 4 basalt-lined elbows or bends, 8-in. I.D.	82,000	29,500
k. Pipeline, spare slurry line to dewatering bins and return waterline (1): Pipeline comprising 34, 40-ft-long sections of flanged steel pipe, 8-in. I.D., schedule 80 carbon steel and 4 hardened elbows or bends, 8-in. I.D.	31,200	11,000
l. Bin, bottom ash dewatering (2): Conical-bottom dewatering bin, 25-ft diameter x 62 ft high, with 2-ft cylindrical section, 18-1/2-ft-high cone, 11,190-ft^3 volume, stainless steel floating decanter and movable drainpipe, stationary decanter in conical section, erected for 16-1/2-ft truck clearance, carbon steel with stainless steel decanter drums, 250-ton capacity	200,000	93,600
m. Settling tank, bottom ash return water (1): 45-ft diameter x 13 ft deep, 154,800 gal, carbon steel, epoxide-coated interior, open top	68,000	38,900

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
n. Surge tank, bottom ash return water (1): Water reservoir, 35-ft diameter x 14 ft deep, 110,000 gal, carbon steel, epoxide-coated interior, open top	48,000	27,300
o. Pump, underflow solids recycle (3): Centrifugal, 250 gpm, 100-ft head, Ni-Hard steel body and impeller, 15 hp (2 operating, 1 spare)	11,300	2,300
p. Pump, dewatering bin sump pit (3): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined (2 operating, 1 spare)	7,200	2,500
3. Water treatment system for recycle water alkalinity control (1):		
a. Storage tank, sulfuric acid, for pH control of water (1): Cylindrical steel tank, 5-ft, 7-in. diameter x 5 ft, 7 in. high, 1,000 gal, flat bottom and closed flat top, carbon steel; all-weather housing	1,900	300
b. Metering pump, sulfuric acid (2): Positive displacement metering pump, 0.01 to 1 gpm, 0 psig, with flow rate controlled by a pH controller, Carpenter 20 alloy or similar corrosion resistance to 93% sulfuric acid, 0.25 hp (1 operating, 1 spare)	1,900	600
c. Agitator, treated water (1): Agitator with 24-in. diameter nickel-chromium blade, 5 hp	2,900	400
Total, Area 13	1,930,500	907,700

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
Area 14--Flue Gas-Handling Modifications[a]		
1. Fan, flue gas (2): Induced draft, 1,031,915 aft^3/min, ΔP = 22 in. H_2O, carbon steel, 5,000-hp motor, fluid drive, double width, double inlet	30,700	400
Total, Area 14	30,700	400
Area 15--Waste Disposal[b]		
1. Landfill site development and construction (1): 80-acre landfill site, 1,533-ft square landfill, 3,559,000-yd^3 volume, 30-yr life, 101 ft high at center, 6,277-ft perimeter ditch to 66,000-yd^3 catchment basin	1,743,400	215,200
2. Wheel loader (1): 5.3-yd^3 bucket, diesel engine	158,600	
3. Dozer (1): Track type with straight blade, 109-hp diesel engine	75,100	
4. Compactor (1): Vibratory sheepsfoot compactor, self-propelled	104,400	
5. Wheel loader (1): 2.6-yd^3 bucket, diesel engine	73,600	
6. Water truck (1): Tandem-axle, 4-rear-wheel-drive tank truck with spray nozzle boom attachment, and pumping system, 1,500-gal fiberglass tank, 130-hp diesel engine	24,400	
7. Service truck (1): Wrecker rig with 500-gal cargo tank for diesel fuel and cargo space for lubricants and other field service items, including tools	27,600	

(Continued)

TABLE 23. (Continued)

Item - Description	Material cost, 1982$	Labor cost, 1982$
8. Onsite trailer for sanitary facilities and break room (1): 12-ft-wide x 30-ft-long mobile home restructured into 2 offices, 1 break room, 1 lavatory; propane gas stove and heater; self-contained portable toilet, potable water supply, and 120-volt electric supply	5,100	
9. Onsite water supply and discharge treatment system (1): Catchment basin pumps, chemical addition tanks and pumps, water supply well, tank, and pumps	20,200	16,500
10. Truck (2): Tandem-axle, 4-rear-wheel-drive dump truck with ash-haul body, 26-yd3 capacity, 56,000-lb suspension, 9 forward speeds, manual transmission, 290-hp diesel engine (1 operating, 1 spare), 75.3% of total truck costs in this area	100,100	
Total, Area 15	2,332,500	231,700

a. Costs shown are additional costs of boiler I.D. fan due to ESP pressure loss.

b. Except as noted, 74.9% of total waste disposal costs is charged to ash removal.

Area 1 - Ammonia Storage and Handling--

This area includes the equipment to receive, store, and inject the ammonia-air mixture into the flue gas. It is identical in design and size to the system in area 1 of case 2 except for the injection and mixing grid configurations which conform to a different inlet duct configuration.

Area 2 - Reactor--

The two reactors are similar in design to the reactors in case 2 but are 49 by 40 feet in cross section (compared with 53 by 37 feet in case 2) and 42 feet high (compared with 43 feet in case 2). The catalyst volume is 27,486 ft^3 and the space velocity is 2,520 hr^{-1}, an 8% smaller catalyst volume and a 7% larger space velocity than case 2 made possible by the absence of fly ash from the flue gas. The only other difference is that the reactors in case 3 do not require ash hopper bottoms.

Area 3 - Flue Gas Handling--

This area consists of the ductwork associated with the NO_x control process, as described in area 3 for case 1, and the incremental increase in the boiler ID fan size to compensate for the 7-inch H_2O pressure drop in the reactors and ducts.

Area 4 - Air Heater Modifications--

The same air heater modifications described in area 4 of cases 1 and 2 are made, but the absence of fly ash allows a lesser increase in element area. Also, water washing frequency is increased since there is no fly ash to aid removal of ammonia salt deposits (through a scouring action). The additional sootblowing and water washing requirements are shown in Table 24.

Area 5 - Waste Disposal--

The waste disposal area description is identical to that of case 1. The spent catalyst volume is smaller because of the smaller volume of catalyst used in case 3.

SO_2 Control

Processing areas 6 through 11 describe the limestone FGD process. In most areas, the process is similar to the description of the FGD process in case 1. The primary differences are in size--related to the different flue gas volumes and SO_2 content--and in the absence of forced oxidation.

Area 6 - Materials Handling--

The equipment necessary to store, retrieve, and supply limestone to the FGD feed preparation area is included in this area. In contrast to the high-sulfur coal case in case 1, which requires 27 ton/hr of limestone feed, this case requires only 4 ton/hr. Because of this, all deliveries are assumed to be by truck, and railcar unloading facilities are not included. The trucks are assumed to dump directly to the storage pile over the reclaim hopper. The reclaim system and the conveying system that transport the limestone to the feed preparation area are the same as those for case 1 except that only one reclaim hopper is used rather than two.

TABLE 24. STEAM SOOTBLOWING AND WATER WASHING REQUIREMENTS

FOR AIR HEATERS OF CASE 3

	Steam sootblowing					
Case	Number of blowers	Cycles/day/ blower	Min/cycle	Lb steam/min	Lb steam/yr	Additional lb steam/yr
Standard without SCR	2	3	20	127	3,492,500	
Modified with SCR	4	3	21	175	10,106,250	6,613,750

	Water washing					
Case	Cycles/yr	Hr/cycle	Gal/ min/heater	Psig	Gal/yr	Additional gal/yr
Standard without SCR	1	8	2,440	150	2,342,400	
Modified with SCR	4	8	2,680	150	10,291,200	7,948,800

Area 7 - Feed Preparation--

The limestone is crushed and wet ball milled to a 40% solids slurry with a particle size of 90% minus 325 mesh in this area. The same equipment as used in area 7 of case 1 is used--including two operating trains and one spare train of crushers and ball mills--but the equipment size is smaller.

Area 8 - Flue Gas Handling--

The flue gas-handling area consists of the inlet plenum that supplies the individual absorber trains, the ductwork of the trains themselves, ID booster fans in each train, and two bypass ducts that serve both for the normal 28% operating bypass and for emergency bypass of 50% of the scrubbed flue gas. The description is the same as the description of area 8 in case 1.

Area 9 - SO_2 Absorption--

This area consists of four trains of spray tower absorbers, one of which is a spare, each 37 by 18.5 feet in plan view and 40 feet high. Except that there is no forced oxidation--the absorbers drain directly to the recirculation tanks--the description is the same as that of area 9 in case 1. The absorbers are designed to scrub 72% of the flue gas at a 90% removal efficiency using three trains (the remaining flue gas is bypassed). The presaturator L/G ratio is 4 gal/kaft^3 and the absorber L/G ratio is 115 gal/kaft^3, the superficial gas velocity is 10 ft/sec, and the stoichiometry is 1.4 mols $CaCO_3$/mol SO_2 + 2HCl absorbed.

Reheat

Because 28% of the flue gas bypasses the absorbers and is recombined with the flue gas in the stack plenum, producing a stack temperature of 175°F, no reheat is required.

Area 10 - Solids Separation--

In this area, the bleedstreams from the absorbers--consisting of an 8% solid slurry, 95% of which is gypsum--are thickened to 40% solids and filtered to 85% solids. Except for the size of the equipment (the thickener is 19 feet in diameter instead of 41 and the filters are 3 feet in diameter and 6 feet long instead of 8 feet and 14 feet, for example), the description of this area is the same as that of area 12 in case 1.

Area 11 - Waste Disposal--

The waste is trucked to the common landfill. Except for the reduction in equipment size and number because of the smaller volume of waste, the description is the same as that for area 13 in case 1.

Particulate Control

Areas 12 to 15 describe the particulate control processes.

Area 12 - Particulate Removal and Storage--

This area consists of the hot-side ESPs and all hoppers associated with ash collection. Two parallel ESPs are used, each 58 feet long, 86 feet wide, and 46 feet high with an SCR of 450 ft^2/kaft^3/min and a removal efficiency of 99.5%. There are 34 double-vee ash hoppers, 10 on the economizer and 24 on the ESPs (there are no ash hoppers on the SCR reactors or air heaters). The description of these hoppers and of the bottom ash hopper is the same as that for area 14 of case 1.

Area 13 - Particulate Transfer--

This area consists of the equipment to remove the ash from the hoppers, dewater the bottom ash, and store the fly ash. The bottom ash system description is the same as that for area 15 of case 1. A pressure pneumatic system similar to that described for area 15 of case 1 is used except that there are fewer hoppers, and thus fewer transfer lines, and the ash storage silos are smaller.

Area 14 - Flue Gas Handling--

This area includes the incremental increase in the boiler ID fan to compensate for the 2-inch H_2O pressure drop in the ESPs and related ductwork and the ductwork connecting the ESPs to the boiler and NO_x control system.

Area 15 - Waste Disposal--

The ash is trucked to the common landfill as described in area 17 of case 1.

RESULTS

Three cases, each composed of processes providing a system for NO_x, SO_2, and particulate matter (including bottom ash) control, serve as the basis of the economic evaluation:

Case 1: 3.5% sulfur coal - SCR reactor - cold-side ESP - limestone FGD

Case 2: 0.7% sulfur coal - SCR reactor - spray dryer FGD - baghouse

Case 3: 0.7% sulfur coal - hot-side ESP - SCR reactor - limestone FGD

Capital investments in 1984 dollars and first-year and levelized annual revenue requirements in 1984 dollars were determined for each process. Both annual revenue requirements contain levelized capital charges; levelized annual revenue requirements also contain levelized operating and maintenance costs. The levelizing factor consists of an annual 10% discount rate and an annual 6% inflation rate. In situations in which the same equipment or function serves more than one process (as in waste disposal and flue gas handling, for example), the costs are prorated between the processes when possible.

The base cases are systems for 500-MW power units, which are described in the systems estimated section. Case variations of these systems were also evaluated to illustrate the effects of different catalyst lives, NO_x reduction levels, and ammonia prices on the cost of NO_x control. In addition, the economics of 200-MW and 1,000-MW systems were determined. These differ from the systems described in the systems estimated section primarily in size and in the number of trains.

It is essential in assessing and comparing the costs of the three systems as a whole, the three processes that compose them, and the component costs of the processes to consider both the type of coal and the interactions created by the particular combinations of processes, both of which have important economic effects. The subbituminous coal contains 80% less sulfur than the bituminous coal and requires removal of 85% less SO_2 to meet the 1979 NSPS. In addition, for the same power output, it produces 17% more flue gas and about 35% less fly ash and bottom ash than the bituminous coal.

These differences in coal properties bear strongly on the types of processes used for SO_2 and particulate control. Up to this time at least, hot-side ESPs and spray dryer FGD would be atypical of high-sulfur coal applications and cold-side ESPs, perhaps less so of low-sulfur coal applications. The combination of individual processes in one emission control system also creates interactions that have important economic effects--the use of spray dryer FGD, for example, combines SO_2 and fly ash control in a way that makes an assessment of some of the separate functions impossible. With the choice of process determined, at least in part, by the type of coal, and the costs associated with individual processes influenced by other processes in the system, economic comparisons on a process-by-process basis must be interpreted with care.

The capital investments and annual revenue requirements for each of the processes and for each of the systems evaluated are shown in the appendix. The same results are summarized in Tables 25 through 28. Detailed cost breakdowns of the three base cases, the case variations, and the energy requirements of the base cases are discussed below.

TABLE 25. SUMMARY OF CAPITAL INVESTMENTS IN M$[a]

	Capital investment, mid-1982 $ M$			
	NO_x	SO_2	Particulate	Total
Base case, 500 MW, 80% NO_x removal				
Case 1	41.9	101.8	42.9	186.6
Case 2	50.1	54.0	62.6	166.7
Case 3	48.1	69.4	53.5	171.0
Case variation, 200 MW, 80% NO_x removal				
Case 1	20.6	58.2	22.6	101.5
Case 2	24.2	31.7	31.4	87.3
Case 3	24.3	41.4	27.8	93.5
Case variation, 1,000 MW, 80% NO_x removal				
Case 1	77.7	175.7	73.3	326.6
Case 2	94.8	97.4	110.7	302.9
Case 3	91.2	121.1	94.6	306.9
Case variation, 500 MW, 90% NO_x removal				
Case 1	48.2	101.9	42.9	193.0
Case 2	55.5	54.0	62.7	172.2
Case 3	53.9	69.4	53.5	176.8

a. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

TABLE 26. SUMMARY OF CAPITAL INVESTMENTS IN \$/kW[a]

	Capital investment, mid-1982 \$ \$/kW			
	NO_x	SO_2	Particulate	Total
Base case, 500 MW, 80% NO_x removal				
Case 1	83.7	203.7	85.8	373.2
Case 2	100.2	108.0	125.3	333.4
Case 3	96.1	138.7	107.1	342.0
Case variation, 200 MW, 80% NO_x removal				
Case 1	103.1	291.0	113.2	507.3
Case 2	121.0	158.3	157.2	436.6
Case 3	121.6	206.9	139.0	467.5
Case variation, 1,000 MW, 80% NO_x removal				
Case 1	77.7	175.7	73.3	326.6
Case 2	94.8	97.4	110.7	302.9
Case 3	91.2	121.1	94.6	306.9
Case variation, 500 MW, 90% NO_x removal				
Case 1	96.4	203.8	85.8	386.0
Case 2	111.0	108.0	125.4	344.3
Case 3	107.8	138.8	107.1	353.6

a. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

BASE CASE COMPARISONS

A breakdown of the capital investment and annual revenue requirements for the base cases illustrates the relative economic importance of each of the processes and the importance of the different cost elements in each of the processes. When equipment or a function serves more than one process, the costs are prorated between the processes if a definable relationship exists: fan capital investment and operating costs are prorated on the basis of the pressure loss in each process; waste disposal costs and waste disposal area land costs are prorated on the basis of waste volume. The baghouse costs are not prorated because the size is largely determined by the flue gas rate.

TABLE 27. SUMMARY OF ANNUAL REVENUE REQUIREMENTS IN M$[a]

	Annual revenue requirements, 1984 $							
	First year				Levelized			
	M$				M$			
	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total
Base case, 500 MW, 80% NO_x removal								
Case 1	21.9	28.8	9.8	60.4	35.8	41.0	12.8	89.7
Case 2	26.5	12.7	14.4	53.6	43.5	16.9	19.0	79.4
Case 3	24.7	18.0	12.1	54.8	40.4	24.9	15.8	81.0
Case variation, 200 MW, 80% NO_x removal								
Case 1	9.7	16.3	5.2	31.2	15.6	23.1	6.9	45.6
Case 2	11.6	7.6	7.7	26.8	18.7	10.1	10.4	39.2
Case 3	11.1	11.0	6.6	28.7	17.8	15.3	8.8	41.9
Case variation, 1,000 MW, 80% NO_x removal								
Case 1	41.5	48.8	16.1	106.4	68.1	69.2	20.9	158.2
Case 2	51.2	22.2	24.5	97.9	84.2	29.2	31.8	145.1
Case 3	47.9	30.3	20.5	98.7	78.4	41.4	26.4	146.3
Case variation, 500 MW, 90% NO_x removal								
Case 1	26.1	28.8	9.8	64.6	42.9	41.0	12.8	96.7
Case 2	30.1	12.7	14.4	57.2	49.5	16.9	19.0	85.4
Case 3	28.6	18.0	12.1	58.6	46.8	24.9	15.8	87.5

a. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

TABLE 28. SUMMARY OF ANNUAL REVENUE REQUIREMENTS IN MILLS/KWH[a]

	Annual revenue requirements, 1984 $							
	Mills/kWh							
	First year				Levelized			
	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total
Base case, 500 MW, 80% NO_x removal								
Case 1	8.0	10.5	3.5	22.0	13.0	14.9	4.7	32.6
Case 2	9.6	4.6	5.2	19.5	15.8	6.2	6.9	28.9
Case 3	9.0	6.5	4.4	19.9	14.7	9.0	5.7	29.5
Case variation, 200 MW, 80% NO_x removal								
Case 1	8.8	14.8	4.7	28.4	14.2	21.0	6.3	41.5
Case 2	10.6	6.9	7.0	24.4	17.0	9.2	9.4	35.7
Case 3	10.1	10.0	6.0	26.1	16.2	13.9	8.0	38.1
Case variation, 1,000 MW, 80% NO_x removal								
Case 1	7.5	8.9	2.9	19.3	12.4	12.6	3.8	28.8
Case 2	9.3	4.0	4.5	17.8	15.3	5.3	5.8	26.4
Case 3	8.7	5.5	3.7	18.0	14.3	7.5	4.8	26.6
Case variation, 500 MW, 90% NO_x removal								
Case 1	9.5	10.5	3.5	23.5	15.6	14.9	4.7	35.2
Case 2	10.9	4.6	5.2	20.8	18.0	6.2	6.9	31.1
Case 3	10.4	6.5	4.4	21.3	17.0	9.0	5.7	31.8

[a] All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

Capital Investment

The base case capital investments by area are shown in Table 29 and are summarized in Figure 6. Those components of indirect capital investment and other capital investment that are determined indirectly, as functions of the total process capital, for example, are listed as a sum identified as "other." For case 1 (3.5% sulfur coal, SCR, limestone FGD, and cold-side ESP), the total capital investment is $187 million, of which NO_x and particulate control accounts for about 22% each and SO_2 control accounts for about 55%. For case 2 (0.7% sulfur coal, SCR, spray dryer FGD, and baghouse), the total capital investment is $167 million with NO_x control accounting for 30%; SO_2 control, 32%; and particulate control, 38%. In this case, however, SO_2 particulate collection is included in the total particulate collection costs rather than SO_2 control costs. For case 3 (0.7% sulfur coal, hot-side ESP, SCR, and limestone FGD), the total capital investment is $171 million with NO_x control accounting for 28%; SO_2 control, 40%; and particulate control, 32%.

Nitrogen Oxides Control--

For NO_x control, the most important capital cost is the initial catalyst charge, which is almost one-third of the total capital investment. Most of the remaining capital costs are for the reactor and the associated internal and external catalyst supports and handling system and for the incremental fan cost and flue gas ductwork associated with flue gas handling. The ammonia storage and injection system constitutes the fourth largest capital cost. The remaining capital costs--air heater modification, waste disposal (of spent catalyst), land, and royalties--are relatively minor. Since the same NO_x control process is used in all three cases, the relationship of the cost elements remains the same in each case.

Most of the capital costs are directly related to the flue gas volume and to a lesser extent, the presence of fly ash. This is particularly true of the major cost areas: catalyst, reactor, and flue gas handling. As a result, the total capital investment for NO_x control in case 1 is $42 million, while for case 2, it is $50 million and for case 3 it is $48 million because of the larger flue gas volume in cases 2 and 3 and the absence of fly ash in case 3. If, however, the costs are expressed in terms of flue gas volume, the capital investments for cases 1, 2, and 3 are 18.6, 17.3, and 16.6 $\$/aft^3/min$, respectively, making the case 1 process the most expensive on this basis.

Because of the low pressure drop in the system, incremental ID fan costs are minor. About 90% of the flue gas-handling costs is for ductwork. The ductwork costs are particularly high in case 3 because of the presence of the hot-side ESP.

Air heater modifications to deal with salt deposits caused by ammonia breakthrough are a minor cost element. Most are associated with the increase in size, the more tightly packed elements, and the use of thicker and more corrosion-resistant elements. Since air heaters are manufactured in incremental sizes, the costs are not directly proportional to the flue gas volumes.

TABLE 29. BASE CASE CAPITAL INVESTMENT COMPARISON

Process capital	Case 1, k$				Case 2, k$				Case 3, k$			
	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total	NO_x	SO_2	Particulate	Total
NH_3 storage and injection	1,314			1,314	1,328			1,328	1,297			1,297
Reactor	7,829			7,829	9,278			9,278	8,453			8,453
Flue gas handling	3,843	11,343	1,311	16,497	4,543	7,374	4,961	16,878	5,386	11,175	4,290	20,851
Air heater	819			819	1,220			1,220	861			861
Materials handling		2,528		2,528		1,132		1,132		1,266		1,266
Feed preparation		4,717		4,717		1,258		1,258		2,363		2,363
SO_2 absorption		20,411		20,411		12,992		12,992		18,070		18,070
Oxidation		2,677		2,677								
Reheat		3,653		3,653								
Solids separation		3,681		3,681						2,265		2,265
Lime particulate recycle						2,140		2,140				
Particulate removal and storage			10,509	10,509			15,446	15,446			14,354	14,354
Particulate transfer			5,636	5,636			6,779	6,779			4,378	4,378
Total process capital, k$	13,805	49,010	17,456	80,271	16,369	24,896	27,186	68,451	15,997	35,139	23,022	74,158
Other Capital Investment												
Waste disposal direct investment	19	4,011	3,344	7,374	34	527	2,749	3,310	30	847	2,628	3,505
Land	10	458	377	845	15	75	326	416	15	113	313	441
Catalyst	12,028			12,028	14,678			14,678	13,455			13,455
Royalty	463			463	563			563	563			563
Other[a]	15,530	48,360	21,710	85,600	18,431	28,478	32,388	79,297	18,001	33,272	27,583	78,856
Total	41,855	101,839	42,887	186,581	50,090	53,976	62,649	166,715	48,061	69,371	53,546	170,978
Total, $/kW[b]	83.7	203.7	85.8	373.2	100.2	108.0	125.3	333.4	96.1	138.7	107.1	342.0

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a. Consists of costs for "services, utilities, and miscellaneous"; all six items of "indirect investment"; "allowance for startup and modifications"; "interest during construction"; and "working capital"; as listed in the appendix tables.

b. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

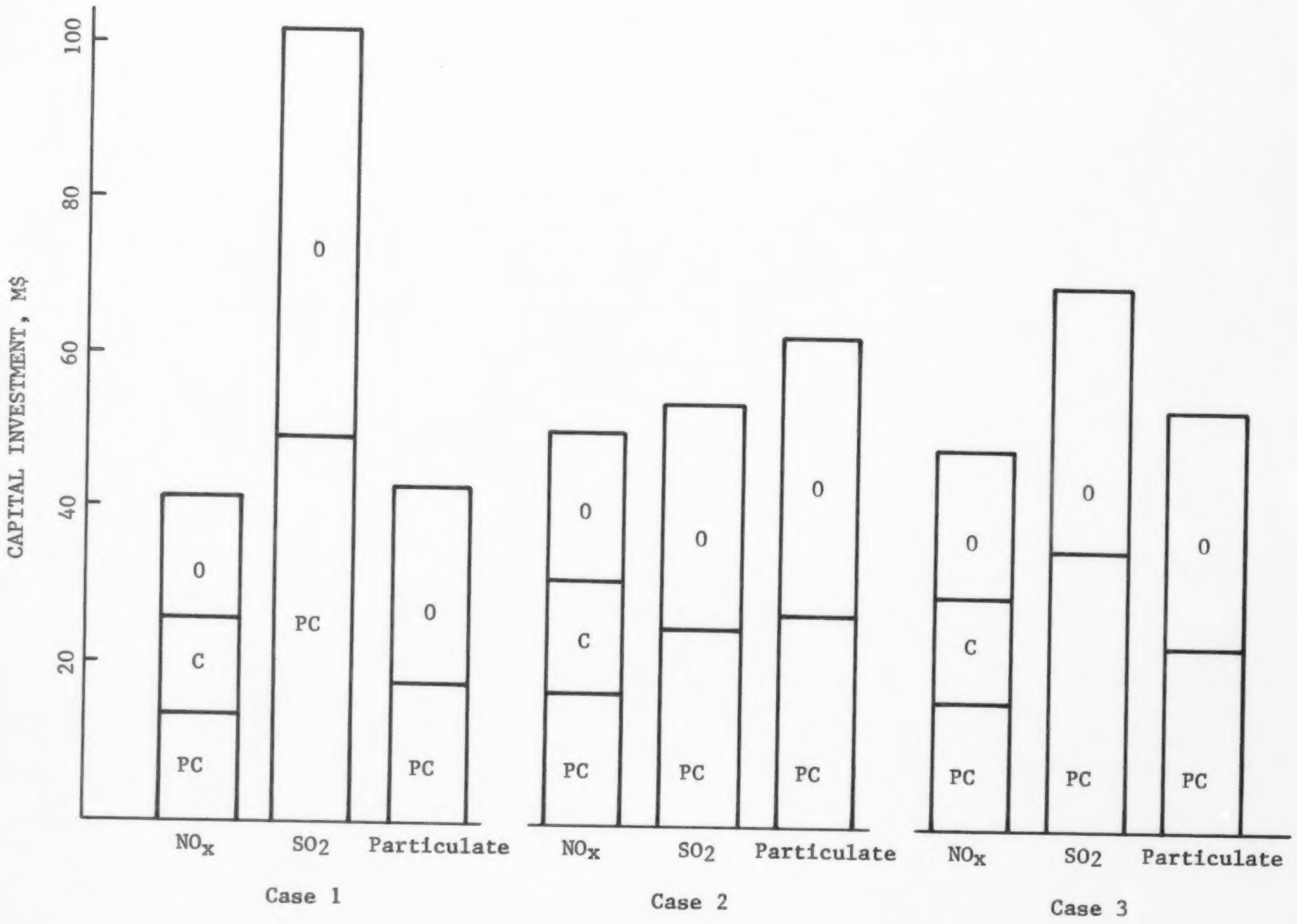

Figure 6. Base case capital investment (PC = process capital, C = initial catalyst, 0 = other).

The air heater modification costs for case 2 are about 50% higher than for case 1, while the same costs for case 3, with the same flue gas volume as case 2 but with no fly ash, are only slightly higher than case 1.

The ammonia storage and injection costs are almost the same for all three cases. The largest cost is for ammonia storage, which is identical for all cases. The only cost differences result from differences in the injection grid, which vary with the flue gas duct size and design.

The costs for process control are included in the processing areas with which they are associated. The total costs for all three cases are about $650,000, about three-fourths in the reactor area (to monitor oxygen, ammonia, and NO_x) and the remainder in the ammonia storage and injection area.

Sulfur Dioxide Control--

The capital investments for SO_2 control are $102 million for case 1, $54 million for case 2, and $69 million for case 3. The capital investment for case 2, however, does not contain the costs for particulate collection, which is an essential function of SO_2 control. Proration of particulate collection cost is not particularly meaningful because the baghouse size is largely determined by the required air-to-cloth ratio--that is, by the flue gas volume rather than the solids rate.

In all three cases, most of the costs are associated with the SO_2 absorption area (the absorbers and the absorbent liquid system or the spray dryers) and the flue gas-handling area (fans and ductwork). These two areas account for 65% of the process equipment costs in case 1 and about 80% of the process equipment costs in cases 2 and 3. The SO_2 absorption area process equipment costs for cases 1 and 3--both of which have limestone scrubbers--do not differ greatly, although case 1 has five trains while case 3 has four trains. The absorbers in case 3 are larger and the L/G ratio is higher, which accounts for the absence of a larger cost difference. In case 2, the cost of the spray dryers themselves is higher than the cost of the absorbers in case 3 ($7.8 million versus $5.1 million) but the spray dryers have no liquid recirculation system while the absorbers in case 3 have about $5 million in tanks, pumps, and piping associated with them.

The flue gas-handling costs for cases 1 and 3 are similar in spite of the additional train in case 1 because of the larger flue gas volume in case 3. The flue gas-handling area costs in case 2 are lower than case 3, largely because of the lower pressure drop in the spray dryers, which reduces the fan costs. Also, the fans in case 2 serve as the booster fans for the baghouse, with the costs prorated between SO_2 control and particulate control, which provides some economy of scale.

The nearly 50% higher capital investment for case 1 as compared with case 3 is almost entirely related to the larger quantities of SO_2 removed. The materials-handling (limestone), feed preparation, and solids separation area costs are roughly two times higher and waste disposal costs are almost five times higher for case 1 than for case 3. In addition, the SO_2 removal requirements in case 1 require full scrubbing; this necessitates steam reheat of the flue gas, which accounts for nearly 8% of the process capital costs. Case 1 also requires forced oxidation; this, however, accounts for only 5% of the process capital cost.

In addition to the lower costs of case 2 in the SO_2 absorption and flue gas-handling areas, it has no solids separation (dewatering) area and lower feed preparation area costs because lime is used, which requires only slaking while the limestone used in case 3 must be crushed and ball milled. Case 2, however, has costs associated with waste recycle that are equivalent to the solids separation area costs of case 3. An accurate comparison of SO_2 control capital investment in cases 2 and 3, however, must include the costs of particulate collection, which is discussed in the following section.

Particulate Control--

The capital investments for particulate control are $43 million for case 1, $63 million for case 2, and $54 million for case 3. The process equipment costs are subdivided into three processing areas: particulate removal and storage, which consists of the ESPs or baghouses (each of the three base cases has two identical units); the hoppers that form their bottoms plus all other ash hoppers on the boiler and SCR reactors; particulate transfer, which consists of the bottom ash sluicing and dewatering system and the fly ash pneumatic conveying system and storage silos; and flue gas handling, which consists of the incremental fan costs and ductwork. Land costs and disposal site development, both prorated from total waste disposal costs, are also included in the total capital investment.

In all three cases, the particulate removal and storage area accounts for about 60% of the total process equipment costs, with the ESPs or baghouses and their hoppers accounting for most of the area cost. The cold-side ESPs of case 1 have an installed cost of $5.9 million and the hot-side ESPs of case 3 have an installed cost of $9.8 million. Most of this difference is a result of the larger flue gas volume in case 3--both in an absolute sense and because the ESPs in case 3 operate at a higher temperature. The ESPs in case 1 have an SCA of 500 but process a flue gas volume of 850,000 aft^3/min. (An SCA of 500 ft^2/kaft^3/min was used for case 1 after determining SCA values ranging from about 450 to over 650 ft^2/kaft^3/min from several references. Some reviewers state that an SCA range of 200 to 250 ft^2/kaft^3/min is adequate to meet the ash removal required by the ESP in case 1. If an ESP designed with an SCA of 250 was used, the investment and revenue requirements for particulate control would be reduced about 15%.) The ESPs in case 3 have an SCA of 450 but process a flue gas volume of 1,447,000 aft^3/min. In addition, the hot-side ESPs of case 3 are constructed to operate at the more rigorous conditions created by the higher temperature. The baghouses have an installed cost of $7.4 million. Much of this cost difference, in comparison to the ESPs, is due to the large size of the baghouses and the corresponding larger and more complex hoppers required. The total volume of the baghouses is 1,656,000 ft^3, compared with 276,000 and 458,000 ft^3 for the cold-side and hot-side ESPs.

Particulate transfer process equipment costs are $5.6 million for case 1, $6.8 million for case 2, and $4.4 million for case 3. Cases 1 and 3 have similar vacuum pneumatic conveying systems and the cost differences are a result of the different quantities of ash. Case 2 has a more complicated pressure-vacuum conveying system, which accounts for most of the cost difference. These costs do not include transportation of the waste to the landfill.

Flue gas-handling costs are $1.3 million for case 1, $5.0 million for case 2, and $4.4 million for case 3. The lower costs for case 1 result from the smaller absolute volume and lower temperature of the flue gas. In addition, the costs are higher for case 3 because of the more complicated and longer duct lengths required for the hot-side ESP. The pressure drops are low for both ESPs and the incremental fan costs are negligible--$21,000 and $31,000. In the case of the baghouse, however, fan costs are a major cost element because of the large pressure drop through the baghouse. For case 2, the incremental fan costs are $2.4 million, almost one-half of the total flue gas-handling costs.

Annual Revenue Requirements

The base case annual revenue requirements are shown in Table 30 and summarized in Figure 7. The first-year annual revenue requirements for case 1 (3.5% sulfur coal, SCR, limestone FGD, and cold-side ESP) are $60 million (22 mills/kWh) with 36% associated with NO_x control, 48% with SO_2 control, and 16% with particulate control. For case 2 (0.7% sulfur coal, SCR, spray dryer FGD, and baghouse), the first-year annual revenue requirements are $54 million (19.5 mills/kWh) with 49% associated with NO_x control, 24% with SO_2 control, and 27% with particulate control. For case 3 (0.7% sulfur coal, hot-side ESP, SCR, and limestone FGD), the first-year annual revenue requirements are $55 million (19.9 mills/kWh) with 45% associated with NO_x control, 33% with SO_2 control, and 22% with particulate control.

The levelized annual revenue requirements are $90 million, $79 million, and $81 million for cases 1, 2, and 3, respectively. For cases 1, 2, and 3, respectively, 40%, 55%, and 50% of the total levelized annual revenue requirements are associated with NO_x control; 46%, 21%, and 31% with SO_2 control; and 14%, 24%, and 19% with particulate control. The percentage of the total associated with NO_x control is higher because of the higher ratio of operating and maintenance costs to capital charges for the SCR process (since the capital charges are levelized in both types of annual revenue requirements, levelizing involves only adjustment of the operating and maintenance costs).

The cost per ton of pollutant removed is presented for the base cases in Table 31 based on each of first-year and levelized annual revenue requirements. A comparison on this basis indicates that NO_x control is significantly less cost effective than SO_2 and particulate control. For example, with first-year annual revenue requirements, the costs in Table 31 range from about 3,500 $/ton to 4,600 $/ton for NO_x control, from about 500 $/ton to over 1,900 $/ton for SO_2 control, and from 60 $/ton to 130 $/ton for particulate control.

TABLE 30. ANNUAL REVENUE REQUIREMENT ELEMENT ANALYSIS FOR BASE CASES

500-MW UNIT WITH 80% NO$_x$ REMOVAL

Direct costs	Case 1				Case 2				Case 3			
	NO$_x$	SO$_2$	Particulate	Total	NO$_x$	SO$_2$	Particulate	Total	NO$_x$	SO$_2$	Particulate	Total
Ammonia	364			364	336			336	336			336
Catalyst	13,889			13,889	16,962			16,962	15,549			15,549
Lime/limestone		1,216		1,216		708		708		186		186
Operating labor and supervision												
Process	66	658	230	954	66	263	296	625	66	594	230	890
Landfill	3	523	436	962	5	83	435	523	4	127	393	524
Steam	51	1,369		1,420	65			65	63			63
Electricity	278	2,146	581	3,005	492	780	964	2,238	391	1,477	993	2,861
Fuel	1	162	135	298	1	18	95	114	1	28	87	116
Maintenance	585	4,276	1,025	5,887	695	1,599	1,811	4,105	679	3,005	1,299	4,983
Analysis	46	104	6	156	46	88	6	140	46	69	6	121
Other	13	27	19	59	17	16	36	69	41	19	36	96
Total direct costs, k$	15,307	10,481	2,432	28,220	18,685	3,555	3,645	25,885	17,176	5,505	3,044	25,725
Indirect costs												
Overheads	421	3,337	1,018	4,776	487	1,220	1,529	3,236	477	2,277	1,157	3,911
Capital charges	6,153	14,970	6,304	27,427	7,363	7,934	9,209	24,506	7,065	10,198	7,871	25,134
Total first-year annual revenue requirements												
k$	21,881	28,788	9,754	60,423	26,535	12,709	14,383	53,627	24,718	17,980	12,072	54,770
Mills/kWh	8.0	10.5	3.5	22.0	9.6	4.6	5.2	19.5	9.0	6.5	4.4	19.9
Levelized annual revenue requirements												
k$	35,816	41,031	12,811	89,658	43,521	16,940	18,967	79,428	40,359	24,875	15,794	81,028
Mills/kWh	13.0	14.9	4.7	32.6	15.8	6.2	6.9	28.9	14.7	9.0	5.7	29.5

a. All values have been rounded; therefore, totals do not necessarily correspond to the sum of the individual values indicated.

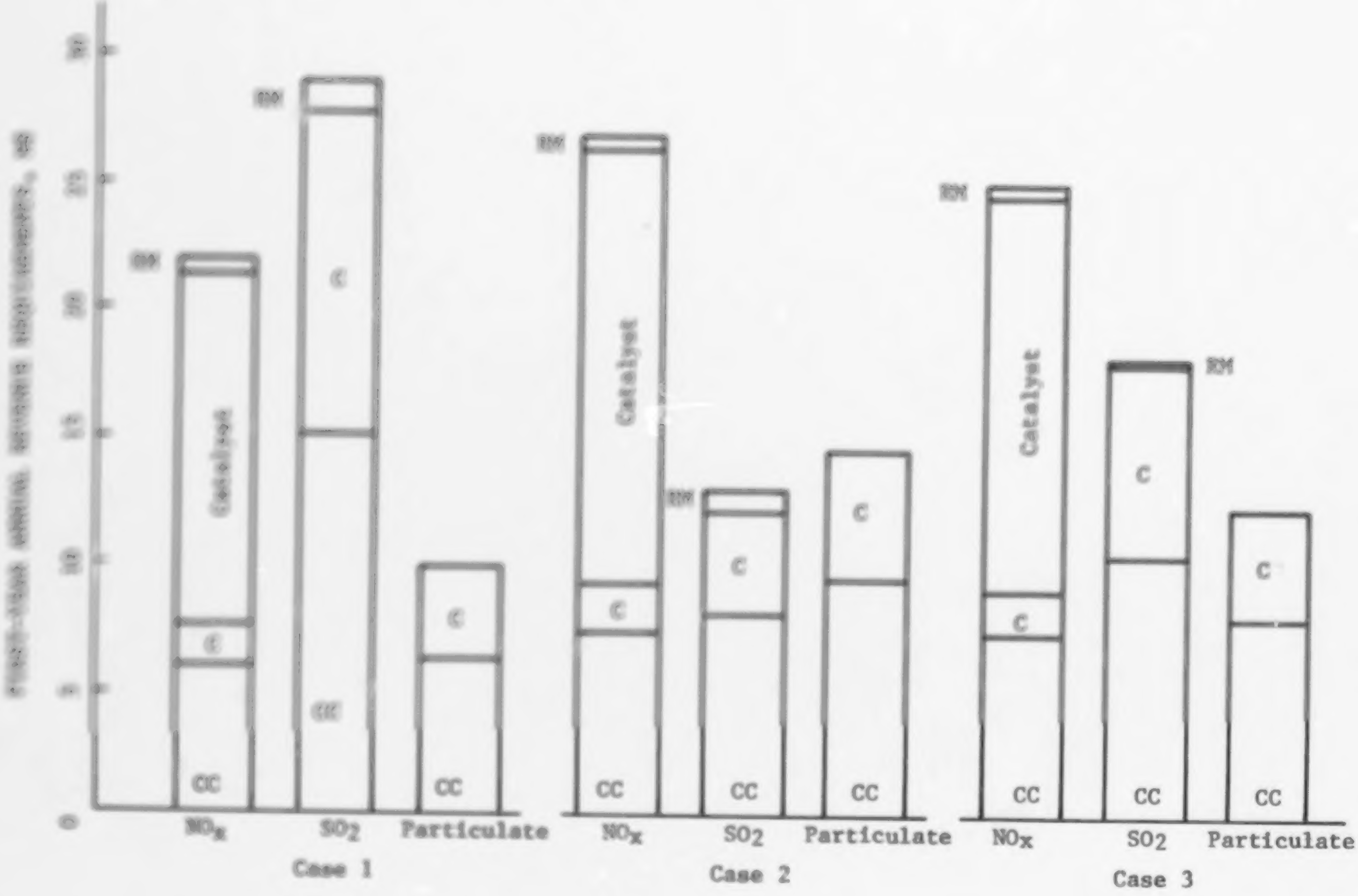

Figure 7. Base case annual revenue requirements (CC = capital charges, C = conversion costs, RM = raw materials).

TABLE 31. COST PER TON OF POLLUTANT REMOVED FOR BASE CASES

500-MW UNIT WITH 80% NO_x REMOVAL

	$/ton, 1984 $					
	First year			Levelized		
	NO_x	SO_2	Particulate	NO_x	SO_2	Particulate
Case 1	3,490	470	60	5,710	670	80
Case 2	4,600	1,370	130	7,540	1,820	170
Case 3	4,280	1,930	110	6,990	2,680	140

Nitrogen Oxides Control--

The first-year annual revenue requirements for the NO_x control processes in cases 1, 2, and 3, respectively, are $22 million, $27 million, and $25 million. In all cases, the catalyst replacement costs are the overwhelmingly dominant cost elements: over 90% of the direct costs and two-thirds of the total annual revenue requirements are for the yearly replacement of catalyst. Except for this cost, the annual revenue requirements are modest, appreciably less than the costs for similar cost categories for SO_2 and particulate control. Maintenance costs are the largest direct cost, followed by ammonia costs and electricity, but the total of these constitute less than 10% of the total direct costs.

Other costs, such as additional steam for air heater sootblowing and catalyst disposal, are negligible. Operating costs for the effects of the process on the air heater operation are for the extra quantities of steam and electricity used in sootblowing, the extra wash water required, the additional chemicals required for wash water treatment, and the extra maintenance (calculated as a percentage of the process capital for air heater modifications). Catalyst disposal operating costs are primarily landfill manpower requirements and maintenance fuel. The sums of these costs as first-year direct operating costs are shown in Table 32 for all three cases evaluated. As can be seen, these operating costs are a very insignificant part of the total annual revenue requirements.

Sulfur Dioxide Control--

The first-year annual revenue requirements for the SO_2 control processes are $29 million, $13 million, and $18 million for cases 1, 2, and 3, respectively. Again, case 2 with the spray dryer does not include costs associated with operation of the baghouse. Excluding capital charges (which are proportional to capital investment) and overheads (which are proportional to the direct costs), the direct costs of the annual revenue requirements reflect appreciably wider differences in operating costs. The direct costs are \$10.5 million, \$3.6 million, and \$5.5 million for cases 1, 2, and 3, respectively. Maintenance costs are the highest element of direct costs in all three cases, followed again in all three cases, by electricity costs. Steam for reheating the flue gas is the third largest direct cost (13% of the total) in case 1, a cost not incurred by cases 2 and 3, which have bypass reheat. Limestone costs are the fourth largest direct cost for case 1, a result of the large quantity of sulfur removed. These costs and the remaining direct costs are all higher than the corresponding costs for cases 2 and 3, a result of the large quantity of SO_2 removed (as opposed to the removal efficiency in the absorber itself). This necessitates full scrubbing, and the accompanying penalty of flue gas reheat requires a large liquid volume and produces a large volume of waste. In contrast, the low-sulfur applications in cases 2 and 3 have much lower direct costs. This is particularly true of the spray dryer process of case 2; with the exception of lime costs, which are 20% of the total direct costs, it has lower direct costs in every category as compared with case 3.

TABLE 32. ADDITIONAL AIR HEATER OPERATION AND CATALYST DISPOSAL COSTS FROM NO_x CONTROL

Case number	Additional air heater operation costs[a], k\$	Catalyst disposal costs, k\$
1	61	5
2	98	7
3	105	6

[a] First-year direct costs, base case conditions.

Particulate Control--

The first-year annual revenue requirements for particulate control are \$10 million, \$14 million, and \$12 million for cases 1, 2, and 3, respectively. The annual revenue requirements for case 2, however, also include the collection of the spray dryer FGD solids. Among the direct costs, maintenance costs are the highest direct cost in all three cases, followed by electricity costs and labor costs. Maintenance costs are highest for case 2, which are about 75% higher than case 1 and 40% higher than case 3. Electricity costs are lowest for case 1 and highest for case 3, while case 2 has only slightly lower electricity costs than case 3. In cases 1 and 3, the electricity costs are primarily for operation of the ESPs; in case 2, the electricity costs are primarily associated with operation of the booster fans. Labor costs do not differ appreciably, although process labor in case 2 is about 25% higher than in cases 1 and 3.

ENERGY REQUIREMENTS

The energy consumptions of the base cases, expressed in Btu equivalents, are shown in Table 33. Almost all of the energy requirements are for electricity to operate the larger boiler ID fans, booster ID fans, the ESPs and absorbent liquid pumps in cases 1 and 3, and the particulate transfer systems. Except in case 1, which has substantial steam requirements for flue gas reheat, steam requirements are minimal, as are the diesel fuel requirements for waste disposal. The total energy requirements range from 4.89% of the boiler capacity for case 1 to 2.31% of the boiler capacity for case 2. The NO_x control energy requirements are the lowest in all three cases. Most are for the incremental electricity consumption of the boiler ID fan that compensates for the relatively small pressure loss in the reactors. For SO_x control, cases 1 and 3 have large electricity requirements because of the FGD booster fans and pumping requirements for the absorbent liquid recirculation systems. These are similar in both cases. Case 1 has higher electricity requirements largely because of the larger feed preparation and waste-handling requirements. The electricity requirements for the spray dryer in case 2 are lower because there is no liquid recirculation system. Particulate control energy requirements in cases 1 and 3 are mostly for ESP electricity, which is substantially lower for the cold-side ESP. In case 2, most of the electricity is for the booster ID fans that compensate for the relatively high pressure drop in the baghouse.

CASE VARIATIONS

Case variations were made to evaluate the effects of power unit size, SCR catalyst life, NO_x reduction level, and ammonia costs.

Power Unit Size Case Variation

The capital investments for 200-MW, 500-MW, and 1,000-MW systems shown in Table 25 are summarized in Figure 8. Compared with the 200-MW systems, the 500-MW systems are 83% to 91% higher and the 1,000-MW systems are 222% to 247% higher in capital investment. In terms of $/kW, the 1,000-MW systems are about one-third less expensive, however, because of the economy of scale. The general relationships of the three cases remain the same at all three power unit sizes. The rate of capital investment increases with increasing power unit size and the rate differs slightly, depending on the processes. The rate of increase is greatest for the NO_x control processes (an increase of 273% to 282% between the 200-MW and 1,000-MW sizes, as compared with 192% to 207% for the SO_2 control processes and 224% to 253% for the particulate control

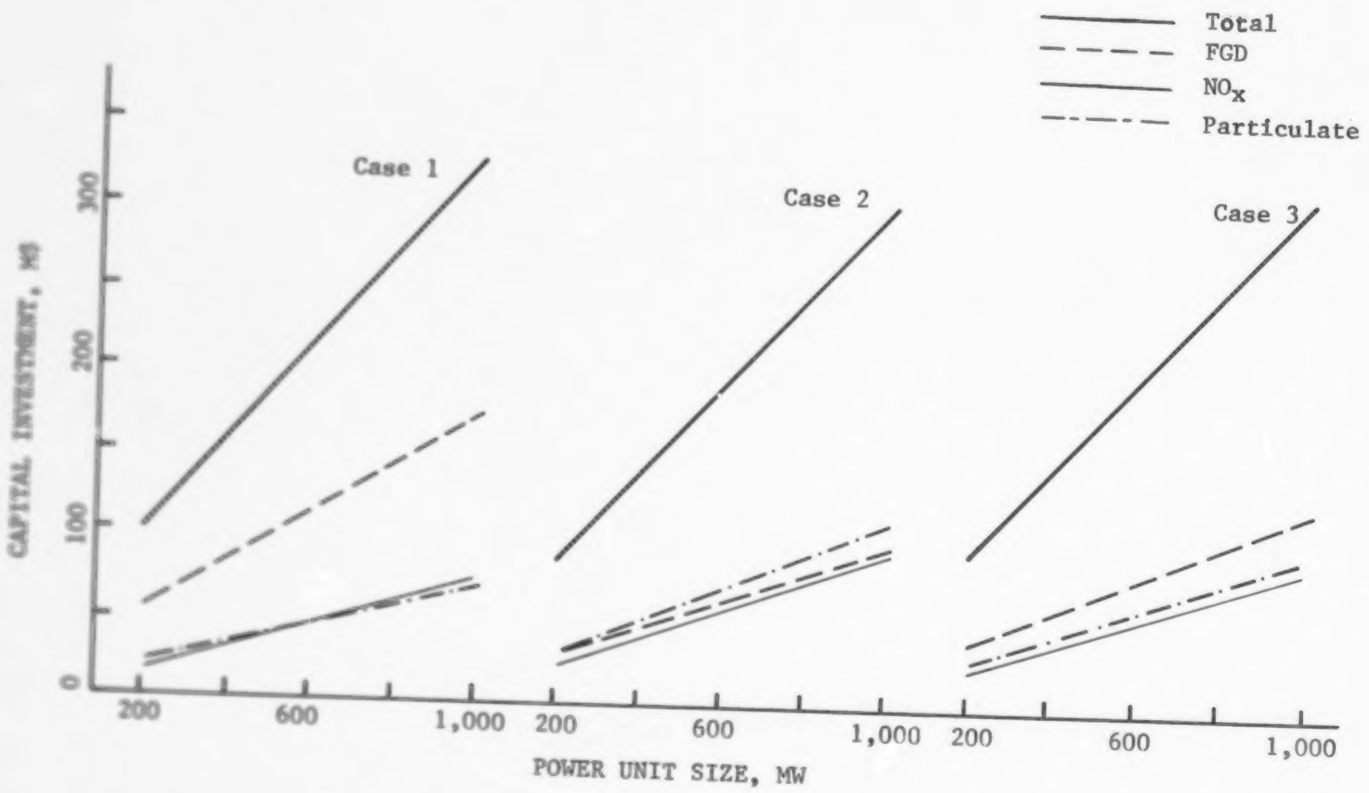

Figure 8. Variation of capital investment with power unit size.

processes) and it is also higher for the spray dryer FGD process and the baghouse than for the limestone FGD process and ESPs. As a result, the rate of capital investment increase with size is greatest for case 2. The rate is, in general, more rapid for costs that depend primarily on the flue gas volume, for which there is less economy of scale; this is most evident in the NO_x control processes in which the flue gas volume is the primary determinant of costs.

TABLE 33. COMPARISON OF BASE CASE ENERGY REQUIREMENTS

Case	Steam, MBtu/hr	Electricity, MBtu/hr	Diesel fuel, MBtu/hr	Percent of power unit, input energy
Case 1[a]				
NO_x	3.15	12.97	0.01	0.34
SO_x	83.79	100.20	2.65	3.93
Particulate	0.00	27.14	2.20	0.62
Total	86.94	140.31	4.86	4.89
Case 2[b]				
NO_x	4.00	25.40	0.02	0.56
SO_x	0.00	40.26	0.30	0.77
Particulate	0.00	49.85	1.55	0.98
Total	4.00	115.51	1.87	2.31
Case 3[b]				
NO_x	3.88	20.18	0.02	0.46
SO_x	0.00	76.20	0.46	1.46
Particulate	0.00	51.22	1.41	1.00
Total	3.88	147.60	1.89	2.92

Note: Does not include energy requirement represented by raw materials.

a. Based on a 500-MW boiler, a gross heat rate of 9,500 Btu/kWh for generation of electricity, and a boiler efficiency of 90% for generation of steam.

b. Based on a 500-MW boiler, a gross heat rate of 10,500 Btu/kWh for generation of electricity, and a boiler efficiency of 90% for generation of steam.

The annual revenue requirements for the same units shown in Table 27 are summarized in Figure 9. Compared with the 200-MW systems, the 500-MW systems are 91% to 100% higher and the 1,000-MW systems are 241% to 265% higher, and there is an approximately one-third reduction in costs in terms of $/kWh. As with capital investment, the annual revenue requirements retain the same general relationships at the three power unit sizes and the rates of increase for the NO_x control processes are higher (328% to 341% between the 200-MW and the 1,000-MW sizes, compared with 175% to 199% for the SO_2 control processes and 210% to 218% for the particulate control processes) and the rates for the spray dryer FGD and baghouse are higher than those of the limestone FGD systems and ESPs.

Two-Year Catalyst Life Case Variation

In the preceding discussions, it has been apparent that the initial and replacement costs of the catalyst are the dominant cost element of both the capital investment and annual revenue requirements for NO_x control and that the annual revenue requirements would be greatly affected by a variation from the assumed 1-year catalyst life. The 1-year life is based on the normal vendor guarantees at the time this project was initiated. Since then, the growing body of experience with SCR processes in coal-fired applications suggests that a longer life--two years or more--is possible. To illustrate the effect of catalyst life on annual revenue requirements, the annual revenue requirements for the three 500-MW base cases were determined for a 2-year catalyst life. The NO_x control capital investment remains unchanged, as do the costs for the SO_2 and particulate control. The only change in NO_x control annual revenue requirements is a reduction in the catalyst cost by 50%-- $7.0 million, $8.5 million, and $7.8 million for cases 1, 2, and 3, respectively. The longer catalyst life reduces the annual revenue requirements of NO_x control by one-third, as shown in Table 34. The annual revenue requirements of the overall systems are reduced by 12% to 16%.

TABLE 34. THE EFFECT OF CATALYST LIFE ON

ANNUAL REVENUE REQUIREMENTS FOR NO_x CONTROL

	Annual catalyst replacement cost, M$		First-year revenue requirements, M$		Levelized annual revenue requirements, M$	
	1-year catalyst life	2-year catalyst life	1-year catalyst life	2-year catalyst life	1-year catalyst life	2-year catalyst life
Case 1	13.9	7.0	21.9	14.9	35.8	22.7
Case 2	17.0	8.5	26.5	18.0	43.5	27.5
Case 3	15.5	7.8	24.7	16.9	40.4	25.7

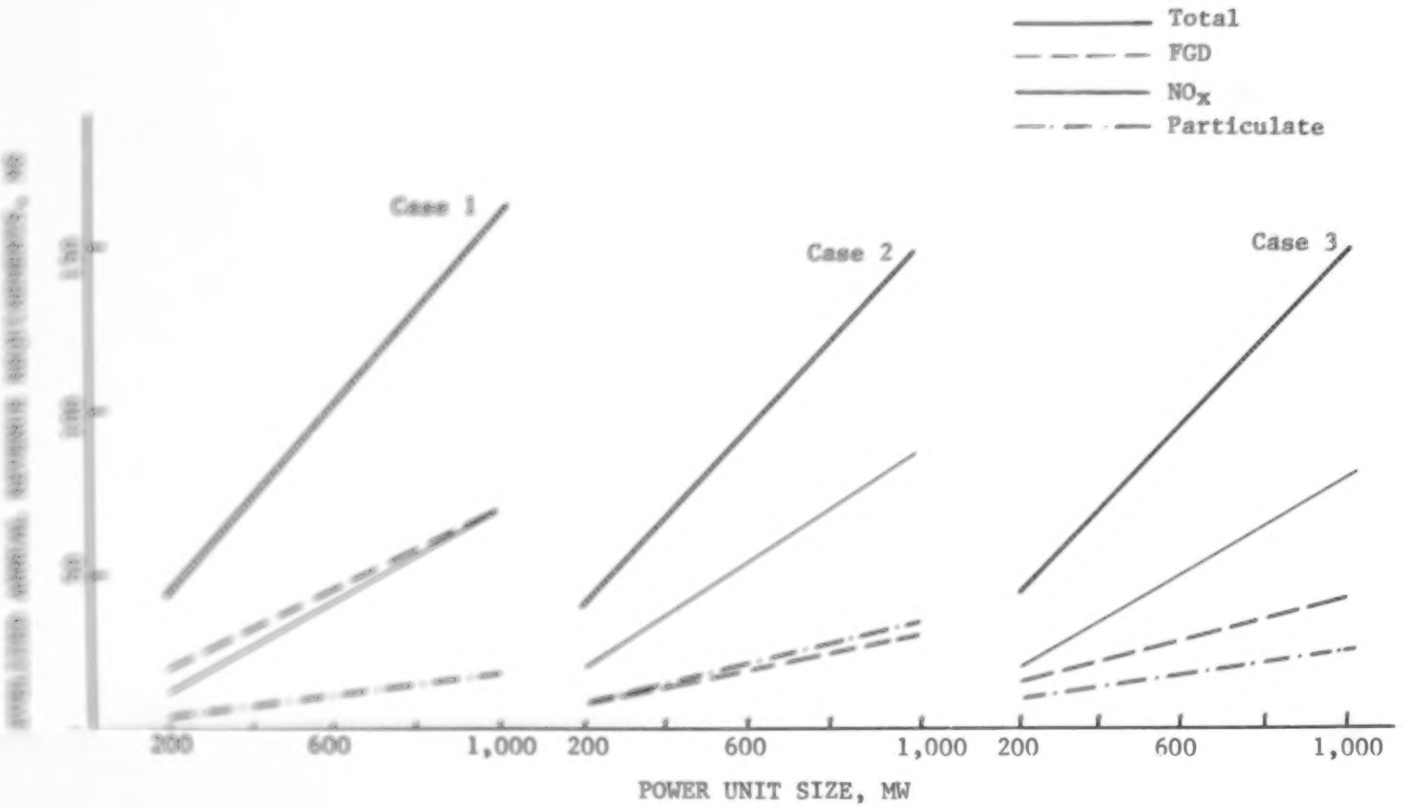

Figure 9. Variation of annual revenue requirements with power unit size.

Ninety Percent Nitrogen Oxide Reduction Case Variation

The SCR processes are capable of very high reduction efficiencies of 90% or greater. However, the reduction rate is dependent on the ammonia addition rate and the catalyst must be increased disproportionately to prevent excessive breakthrough of ammonia, as well as to obtain the higher reduction efficiency. To evaluate the economic effects of a 90% reduction in NO_x, as compared with the 80% used in the other evaluations, the economics of the three 500-MW cases were determined with 90% NO_x reduction. The results are summarized in Tables 25 and 27 and the annual revenue requirements for 80% and 90% reduction are compared in Figure 10.

The primary differences from the base case conditions are an NH_3:NO_x ratio of 0.91:1.0 instead of 0.81:1.0, a 12% increase, and an increase in catalyst (based on vendor recommendations) of 22.5% for case 1, 15.0% for case 2, and 18.0% for case 3. There are also slight increases in the flue gas volume because of the additional ammonia and air, which have slight effects on the SO_2 and particulate control processes.

The capital investments of the NO_x control processes are increased 11% to 15% and the total for the three systems by 3% to 4%, all of which is a result of the increase in NO_x reduction. Most of the increase in NO_x control capital investment is a result of the increase in the reactor size; the increase is proportional to the increase in catalyst and the increase in reactor costs is also proportional to this increase. These account for most of the capital investment increases.

The first-year annual revenue requirements for the NO_x process are increased 19%, 14%, and 16% for cases 1, 2, and 3, respectively. Most of the increase is the result of the additional catalyst replacement cost and the increase in capital charges. The ammonia costs are increased 12%, which has little effect on the total annual revenue requirements. The annual revenue requirements of the other processes are affected little. The effect on the annual revenue requirements of the overall system of increasing the NO_x reduction from 80% to 90% is an increase of 7% in all three cases.

Ammonia Price Case Variation

The possibility of increases in the prices of high-priced raw materials is an economic concern in some emission control processes. The SCR NO_x control process uses such a raw material--ammonia, for which a price of 155 $/ton is used in this study. In this case, however, the costs associated with the ammonia are a minor element of the annual revenue requirements--less than 2% of the total in the 500-MW base cases. Under these conditions, changes in the price of ammonia would have little effect on the overall cost of the process, as shown in Table 35, which shows the annual revenue requirements for the NO_x control processes at an ammonia price of 310 $/ton. The annual revenue requirements are increased 1.5% to 1.9%.

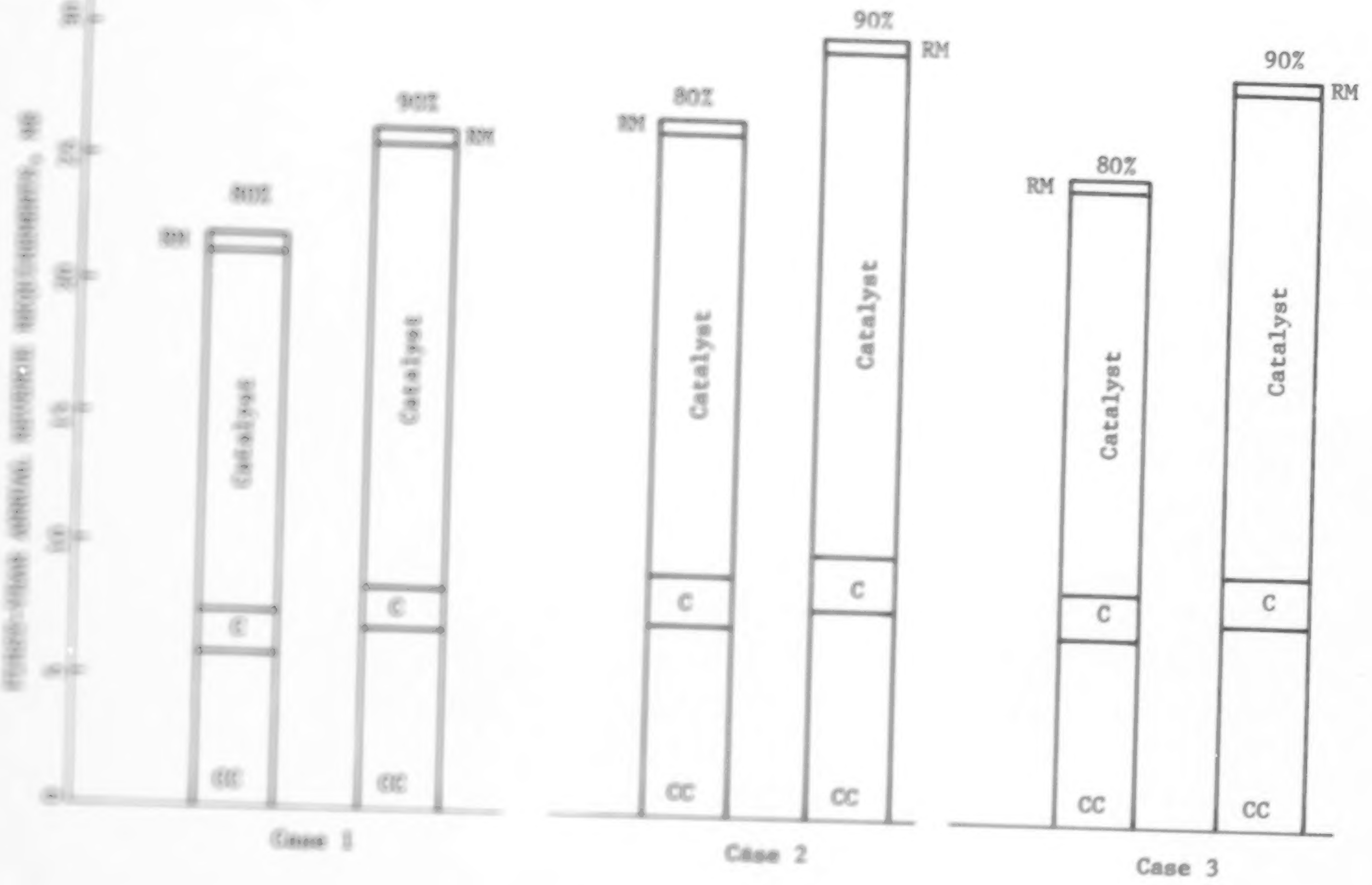

Figure [illegible]. Annual revenue requirements for 80% and 90% NO_x reduction (CC = capital charges, C = conversion costs, RM = raw materials).

TABLE 35. SENSITIVITY OF NO_x CONTROL TO AMMONIA PRICE

Case number	Levelized annual revenue requirements				Percent change in revenue requirement due to increased NH_3 costs
	NH_3 at $155/ton		NH_3 at 310 $/ton		
	M$/yr	Mills/kWh	M$/yr	Mills/kWh	
1	35,816	13.0	36,507	13.3	1.9
2	43,521	15.8	44,159	16.1	1.5
3	40,359	14.7	40,997	14.9	1.6

CONCLUSIONS

The total costs for case 1, based on 3.5% sulfur coal, and cases 2 and 3, based on 0.7% sulfur coal, differ less than 15% in capital investment and annual revenue requirements in spite of the differing coals and control processes. This is a result in part of offsetting differences--the much higher SO_2 control costs for case 1 are offset by lower fly ash control costs and a smaller flue gas volume. The costs for the two low-sulfur coal cases, one with a spray dryer FGD system and baghouse and the other with limestone FGD and a hot-side ESP, differ only marginally in cost. In the two low-sulfur coal cases, the low spray dryer FGD costs and the advantage of combined particulate collection are offset by the higher NO_x control costs and higher baghouse costs. When only the SO_2 and fly ash control costs are compared, the spray dryer-baghouse case is 5% lower in capital investment and 12% lower in annual revenue requirements than the hot-side ESP limestone FGD case.

The combined emission control processes increase the power plant capital investment by about 35% on the average, of which the NO_x portion is about one-third. Based on levelized annual revenue requirements, the average increase in the cost of power is about 45%, of which the NO_x portion is about one-half.

The energy requirements of 2% to 5% of the boiler input energy are mostly for SO_2 and particulate control. For the cases with limestone FGD, SO_2 control has the highest energy requirements.

The use of flue gas treatment for NO_x control, such as the SCR process in this study, would add significantly to emission control costs. An SCR process for a 500-MW power plant would have a capital investment of 80 to 100 $/kW and annual revenue requirements of 8 to 9 mills/kWh. The high cost is largely associated with the catalyst replacement cost, which accounts for 90% of the direct costs in annual revenue requirements. A 2-year catalyst life reduces the annual revenue requirements by over one-third, however, so the costs for NO_x control in this study, which are based on a 1-year life, could be substantially reduced if extended catalyst lives are attained.

Other than catalyst life, the main factor affecting NO_x control costs is the flue gas volume which determines the fan and ductwork costs and the catalyst volume. Increasing the NO_x reduction efficiency from 80% to 90% increases the costs by 10% to 20%, again because of the larger catalyst volume needed. Ammonia costs have almost no effect on overall costs; doubling the price of ammonia increases the annual revenue requirements by about 2%.

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Although the costs of NO_x control are in the same general range as those for SO_2 and fly ash control, if the processes are compared on the basis of the pounds of pollutants reduced, the costs for NO_x control are 2 to 10 times greater than for SO_2 control and 40 to 60 times greater than for ash control.

In SO_2 control, the major costs are associated with the absorption area and flue gas handling (ductwork and fans). These costs do not differ greatly among the three cases because of offsetting differences--a larger cost for liquid circulation in the high-sulfur coal case but a larger flue gas volume in the low-sulfur coal cases, which requires larger equipment and has larger fan costs. The higher costs for the high-sulfur coal case are in large part the result of the much larger quantity of sulfur removed: the materials-handling, waste-handling, and disposal costs are two to five times higher for the high-sulfur coal case than for the low-sulfur coal case with limestone FGD.

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APPENDIX A

TABLE A-1. CASE 1, 200-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	745
Reactor	3,373
Flue gas handling	2,441
Air heater	478
Total process capital	7,037
Services, utilities, and miscellaneous	422
Total direct investment excluding waste disposal	7,459
Waste disposal	10
Total direct investment	7,469
Indirect Investment	
Engineering design and supervision	597
Architect and engineering contractor	224
Construction expense	1,343
Contractor fees	448
Contingency	2,014
Waste disposal indirect investment	4
Total fixed investment	12,099
Other Capital Investment	
Allowance for startup and modifications	1,209
Interest during construction	1,887
Royalties	192
Land	5
Working capital	343
Catalyst	4,893
Total capital investment	20,628
$/kW	103.1

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-2. CASE 1, 200-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	959 tons	155	149
Catalyst	240 tons	23,558	5,654
Sodium hydroxide	16 tons	300	5
H_2SO_4 (100% equivalent)	8 tons	65	1
Total raw material cost			5,809
Conversion costs			
Operating labor and supervision			
Process	2,190 man-hr	15.00	33
Landfill	77 man-hr	21.00	2
Utilities			
Steam	6,358 MBtu	3.30	21
Process water	1,074 kgal	0.14	0
Electricity	3,068,548 kWh	0.037	114
Diesel fuel	201 gal	1.60	0
Maintenance			
Labor and material			373
Analysis	1,752 man-hr	21.00	37
Total conversion costs			580
Total direct costs			6,389
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			267
Total first-year operating and maintenance costs			6,656
Levelized capital charges (14.7% of total capital investment)			3,032
Total first-year annual revenue requirements			9,688
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			12,553
Levelized capital charges (14.7% of total capital investment)			3,032
Total levelized annual revenue requirements			15,585

	M$	Mills/kWh
First-year annual revenue requirements	9.7	8.8
Levelized annual revenue requirements	15.6	14.2

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-3. CASE 1, 200-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	1,283
Feed preparation	2,748
Flue gas handling	6,013
SO_2 absorption	10,880
Oxidation	1,540
Reheat	1,899
Solids separation	2,781
Total process capital	27,144
Services, utilities, and miscellaneous	1,629
Total direct investment excluding waste disposal	28,773
Waste disposal	2,069
Total direct investment	30,842
Indirect Investment	
Engineering design and supervision	2,302
Architect and engineering contractor	863
Construction expense	5,179
Contractor fees	1,726
Contingency	3,884
Waste disposal indirect investment	756
Total fixed investment	45,552
Other Capital Investment	
Allowance for startup and modifications	3,418
Interest during construction	7,106
Land	256
Working capital	1,871
Total capital investment	58,203
$/kW	291.0

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level. forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-4. CASE 1, 200-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Limestone	58,400 tons	8.50	496
Total raw material cost			496
Conversion costs			
Operating labor and supervision			
Process	33,470 man-hr	15.00	502
Landfill	15,825 man-hr	21.00	332
Utilities			
Steam	169,394 MBtu	3.30	559
Process water	79,170 kgal	0.14	11
Electricity	24,564,866 kWh	0.037	909
Diesel fuel	41,073 gal	1.60	66
Maintenance			
Labor and material			2,652
Analysis	3,300 man-hr	21.00	69
Total conversion costs			5,100
Total direct costs			5,596
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,133
Total first-year operating and maintenance costs			7,729
Levelized capital charges (14.7% of total capital investment)			8,556
Total first-year annual revenue requirements			16,285
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			14,577
Levelized capital charges (14.7% of total capital investment)			8,556
Total levelized annual revenue requirements			23,133

	M$	Mills/kWh
First-year annual revenue requirements	16.3	14.8
Levelized annual revenue requirements	23.1	21.0

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-5. CASE 1, 200-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Particulate removal and storage	4,679
Particulate transfer	3,356
Flue gas handling	833
Total process capital	8,868
Services, utilities, and miscellaneous	532
Total direct investment excluding waste disposal	9,400
Waste disposal	1,730
Total direct investment	11,130
Indirect Investment	
Engineering design and supervision	752
Architect and engineering contractor	282
Construction expense	1,692
Contractor fees	564
Contingency	2,538
Waste disposal indirect investment	631
Total fixed investment	17,589
Other Capital Investment	
Allowance for startup and modifications	1,523
Interest during construction	2,744
Land	207
Working capital	570
Total capital investment	22,633
$/kW	113.2

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-6. CASE 1, 200-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	112 tons	65	7
Total raw material cost			7
Conversion costs			
Operating labor and supervision			
Process	6,570 man-hr	15.00	99
Landfill	13,218 man-hr	21.00	278
Utilities			
Process water	3,083 kgal	0.14	0
Electricity	6,435,402 kWh	0.037	238
Diesel fuel	34,308 gal	1.60	55
Maintenance			
Labor and material			616
Analysis	200 man-hr	21.00	4
Total conversion costs			1,290
Total direct costs			1,297
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			598
Total first-year operating and maintenance costs			1,895
Levelized capital charges (14.7% of total capital investment)			3,327
Total first-year annual revenue requirements			5,222
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			3,574
Levelized capital charges (14.7% of total capital investment)			3,327
Total levelized annual revenue requirements			6,901

	M$	Mills/kWh
First-year annual revenue requirements	5.2	4.7
Levelized annual revenue requirements	6.9	6.3

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-7. CASE 1, 200-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	7,037
SO_2 removal areas	27,144
Particulate removal areas	8,868
Total process capital	43,049
Services, utilities, and miscellaneous	2,583
Total direct investment excluding waste disposal	45,632
Waste disposal	3,809
Total direct investment	49,441
Indirect Investment	
Engineering design and supervision	3,651
Architect and engineering contractor	1,369
Construction expense	8,214
Contractor fees	2,738
Contingency	8,436
Waste disposal indirect investment	1,391
Total fixed investment	75,240
Other Capital Investment	
Allowance for startup and modifications	6,150
Interest during construction	11,737
Royalties	192
Land	468
Working capital	2,784
Catalyst	4,893
Total capital investment	101,464
$/kW	507.3

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-8. CASE 1, 200-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	959 tons	155	149
Catalyst	240 tons	23,558	5,654
Sodium hydroxide	16 tons	300	5
H_2SO_4 (100% equivalent)	120 tons	65	8
Limestone	58,400 tons	8.50	496
Total raw material cost			6,312
Conversion costs			
Operating labor and supervision			
Process	42,230 man-hr	15.00	634
Landfill	29,120 man-hr	21.00	612
Utilities			
Steam	175,752 MBtu	3.30	580
Process water	83,326 kgal	0.14	11
Electricity	34,068,816 kWh	0.037	1,261
Diesel fuel	75,582 gal	1.60	121
Maintenance			
Labor and material			3,641
Analysis	5,252 man-hr	21.00	110
Total conversion costs			6,970
Total direct costs			13,282
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,998
Total first-year operating and maintenance costs			16,280
Levelized capital charges (14.7% of total capital investment)			14,915
Total first-year annual revenue requirements			31,195
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			30,704
Levelized capital charges (14.7% of total capital investment)			14,915
Total levelized annual revenue requirements			45,619

	M$	Mills/kWh
First-year annual revenue requirements	31.2	28.4
Levelized annual revenue requirements	45.6	41.5

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-9. CASE 1, 500-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	1,314
Reactor	7,829
Flue gas handling	3,843
Air heater	819
Total process capital	13,805
Services, utilities, and miscellaneous	828
Total direct investment excluding waste disposal	14,633
Waste disposal	19
Total direct investment	14,652
Indirect Investment	
Engineering design and supervision	1,024
Architect and engineering contractor	293
Construction expense	2,341
Contractor fees	732
Contingency	3,805
Waste disposal indirect investment	7
Total fixed investment	22,854
Other Capital Investment	
Allowance for startup and modifications	2,283
Interest during construction	3,565
Royalties	463
Land	10
Working capital	652
Catalyst	12,028
Total capital investment	41,855
$/kW	83.7

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-10. CASE 1, 500-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,350 tons	155	364
Catalyst	590 tons	23,558	13,899
Sodium hydroxide	39 tons	300	12
H_2SO_4 (100% equivalent)	20 tons	65	1
Total raw material cost			14,276
Conversion costs			
Operating labor and supervision			
Process	4,380 man-hr	15.00	66
Landfill	122 man-hr	21.00	3
Utilities			
Steam	15,572 MBtu	3.30	51
Process water	2,629 kgal	0.14	0
Electricity	7,508,752 kWh	0.037	278
Diesel fuel	498 gal	1.60	1
Maintenance			
Labor and material			586
Analysis	2,190 man-hr	21.00	46
Total conversion costs			1,031
Total direct costs			15,307
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			421
Total first-year operating and maintenance costs			15,728
Levelized capital charges (14.7% of total capital investment)			6,153
Total first-year annual revenue requirements			21,881
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			29,663
Levelized capital charges (14.7% of total capital investment)			6,153
Total levelized annual revenue requirements			35,816

	M$	Mills/kWh
First-year annual revenue requirements	21.9	8.0
Levelized annual revenue requirements	35.8	13.0

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-11. CASE 1, 500-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	2,528
Feed preparation	4,717
Flue gas handling	11,343
SO_2 absorption	20,411
Oxidation	2,677
Reheat	3,653
Solids separation	3,681
Total process capital	49,010
Services, utilities, and miscellaneous	2,941
Total direct investment excluding waste disposal	51,951
Waste disposal	4,011
Total direct investment	55,962
Indirect Investment	
Engineering design and supervision	3,637
Architect and engineering contractor	1,039
Construction expense	8,312
Contractor fees	2,598
Contingency	6,754
Waste disposal indirect investment	1,355
Total fixed investment	79,657
Other Capital Investment	
Allowance for startup and modifications	5,943
Interest during construction	12,426
Land	458
Working capital	3,355
Total capital investment	101,839
$/kW	203.7

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-12. CASE 1, 500-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Limestone	143,030 tons	8.50	1,216
Total raw material cost			1,216
Conversion costs			
Operating labor and supervision			
Process	43,860 man-hr	15.00	658
Landfill	24,896 man-hr	21.00	523
Utilities			
Steam	414,758 MBtu	3.30	1,369
Process water	193,864 kgal	0.14	27
Electricity	58,011,017 kWh	0.037	2,146
Diesel fuel	101,256 gal	1.60	162
Maintenance			
Labor and material			4,276
Analysis	4,940 man-hr	21.00	104
Total conversion costs			9,265
Total direct costs			10,481
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,337
Total first-year operating and maintenance costs			13,818
Levelized capital charges (14.7% of total capital investment)			14,970
Total first-year annual revenue requirements			28,788
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			26,061
Levelized capital charges (14.7% of total capital investment)			14,970
Total levelized annual revenue requirements			41,031

	M$	Mills/kWh
First-year annual revenue requirements	28.8	10.5
Levelized annual revenue requirements	41.0	14.9

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-13. CASE 1, 500-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Particulate removal and storage	10,509
Particulate transfer	5,636
Flue gas handling	1,311
Total process capital	17,456
Services, utilities, and miscellaneous	1,047
Total direct investment excluding waste disposal	18,503
Waste disposal	3,344
Total direct investment	21,847
Indirect Investment	
Engineering design and supervision	1,295
Architect and engineering contractor	370
Construction expense	2,960
Contractor fees	925
Contingency	4,811
Waste disposal indirect investment	1,129
Total fixed investment	33,337
Other Capital Investment	
Allowance for startup and modifications	2,886
Interest during construction	5,201
Land	377
Working capital	1,086
Total capital investment	42,887
$/kW	85.8

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-14. CASE 1, 500-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	275 tons	65	18
Total raw material cost			18
Conversion costs			
Operating labor and supervision			
Process	15,330 man-hr	15.00	230
Landfill	20,742 man-hr	21.00	436
Utilities			
Process water	7,549 kgal	0.14	1
Electricity	15,715,143 kWh	0.037	581
Diesel fuel	84,360 gal	1.60	135
Maintenance			
Labor and material			1,025
Analysis	300 man-hr	21.00	6
Total conversion costs			2,414
Total direct costs			2,432
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,018
Total first-year operating and maintenance costs			3,450
Levelized capital charges (14.7% of total capital investment)			6,304
Total first-year annual revenue requirements			9,754
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			6,507
Levelized capital charges (14.7% of total capital investment)			6,304
Total levelized annual revenue requirements			12,811

	M$	Mills/kWh
First-year annual revenue requirements	9.8	3.5
Levelized annual revenue requirements	12.8	4.7

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-15. CASE 1, 500-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	13,805
SO_2 removal areas	49,010
Particulate removal areas	17,456
Total process capital	80,271
Services, utilities, and miscellaneous	4,816
Total direct investment excluding waste disposal	85,087
Waste disposal	7,374
Total direct investment	92,461
Indirect Investment	
Engineering design and supervision	5,956
Architect and engineering contractor	1,702
Construction expense	13,613
Contractor fees	4,255
Contingency	15,370
Waste disposal indirect investment	2,491
Total fixed investment	135,848
Other Capital Investment	
Allowance for startup and modifications	11,112
Interest during construction	21,192
Royalties	463
Land	845
Working capital	5,093
Catalyst	12,028
Total capital investment	186,581
$/kW	373.2

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-16. CASE 1, 500-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,350 tons	155	364
Catalyst	590 tons	23,558	13,899
Sodium hydroxide	39 tons	300	12
H_2SO_4 (100% equivalent)	295 tons	65	19
Limestone	143,030 tons	8.50	1,216
Total raw material cost			15,510
Conversion costs			
Operating labor and supervision			
Process	63,570 man-hr	15.00	954
Landfill	45,760 man-hr	21.00	962
Utilities			
Steam	430,330 MBtu	3.30	1,420
Process water	204,042 kgal	0.14	28
Electricity	81,234,912 kWh	0.037	3,005
Diesel fuel	186,114 gal	1.60	298
Maintenance			
Labor and material			5,887
Analysis	7,430 man-hr	21.00	156
Total conversion costs			12,710
Total direct costs			28,220
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,776
Total first-year operating and maintenance costs			32,996
Levelized capital charges (14.7% of total capital investment)			27,427
Total first-year annual revenue requirements			60,423
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			62,231
Levelized capital charges (14.7% of total capital investment)			27,427
Total levelized annual revenue requirements			89,658

	M$	Mills/kWh
First-year annual revenue requirements	60.4	22.0
Levelized annual revenue requirements	89.7	32.6

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-17. CASE 1, 500-MW, NO_x REMOVAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	1,414
Reactor	9,469
Flue gas handling	3,845
Air heater	819
Total process capital	15,547
Services, utilities, and miscellaneous	933
Total direct investment excluding waste disposal	16,480
Waste disposal	24
Total direct investment	16,504
Indirect Investment	
Engineering design and supervision	1,154
Architect and engineering contractor	330
Construction expense	2,637
Contractor fees	824
Contingency	4,285
Waste disposal indirect investment	8
Total fixed investment	25,742
Other Capital Investment	
Allowance for startup and modifications	2,571
Interest during construction	4,016
Royalties	463
Land	13
Working capital	728
Catalyst	14,678
Total capital investment	48,211
$/kW	96.4

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-18. CASE 1, 500-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,640 tons	155	409
Catalyst	720 tons	23,558	16,962
Sodium hydroxide	39 tons	300	12
H_2SO_4 (100% equivalent)	20 tons	65	1
Total raw material cost			17,384
Conversion costs			
Operating labor and supervision			
Process	4,380 man-hr	15.00	66
Landfill	149 man-hr	21.00	3
Utilities			
Steam	16,941 MBtu	3.30	56
Process water	2,635 kgal	0.14	0
Electricity	7,654,216 kWh	0.037	283
Diesel fuel	606 gal	1.60	1
Maintenance			
Labor and material			660
Analysis	2,190 man-hr	21.00	46
Total conversion costs			1,115
Total direct costs			18,499
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			465
Total first-year operating and maintenance costs			18,964
Levelized capital charges (14.7% of total capital investment)			7,087
Total first-year annual revenue requirements			26,051
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			35,766
Levelized capital charges (14.7% of total capital investment)			7,087
Total levelized annual revenue requirements			42,853

	M$	Mills/kWh
First-year annual revenue requirements	26.1	9.5
Levelized annual revenue requirements	42.9	15.6

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-19. CASE 1, 500-MW, SO_2 REMOVAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
Materials handling	2,528
Feed preparation	4,717
Flue gas handling	11,354
SO_2 absorption	20,413
Oxidation	2,689
Reheat	3,653
Solids separation	3,681
Total process capital	49,035
Services, utilities, and miscellaneous	2,942
Total direct investment excluding waste disposal	51,977
Waste disposal	4,011
Total direct investment	55,988
Indirect Investment	
Engineering design and supervision	3,638
Architect and engineering contractor	1,040
Construction expense	8,316
Contractor fees	2,599
Contingency	6,757
Waste disposal indirect investment	1,355
Total fixed investment	79,693
Other Capital Investment	
Allowance for startup and modifications	5,946
Interest during construction	12,432
Land	458
Working capital	3,357
Total capital investment	101,886
$/kW	203.8

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-20. CASE 1, 500-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Limestone	143,058 tons	8.50	1,216
Total raw material cost			1,216
Conversion costs			
Operating labor and supervision			
Process	43,860 man-hr	15.00	658
Landfill	24,896 man-hr	21.00	523
Utilities			
Steam	415,182 MBtu	3.30	1,370
Process water	194,024 kgal	0.14	27
Electricity	58,055,993 kWh	0.037	2,148
Diesel fuel	101,256 gal	1.60	162
Maintenance			
Labor and material			4,278
Analysis	4,940 man-hr	21.00	104
Total conversion costs			9,270
Total direct costs			10,486
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,338
Total first-year operating and maintenance costs			13,824
Levelized capital charges (14.7% of total capital investment)			14,977
Total first-year annual revenue requirements			28,801
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			26,072
Levelized capital charges (14.7% of total capital investment)			14,977
Total levelized annual revenue requirements			41,049

	M$	Mills/kWh
First-year annual revenue requirements	28.8	10.5
Levelized annual revenue requirements	41.0	14.9

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-21. CASE 1, 500-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
Particulate removal and storage	10,519
Particulate transfer	5,636
Flue gas handling	1,312
Total process capital	17,467
Services, utilities, and miscellaneous	1,048
Total direct investment excluding waste disposal	18,515
Waste disposal	3,344
Total direct investment	21,859
Indirect Investment	
Engineering design and supervision	1,296
Architect and engineering contractor	370
Construction expense	2,962
Contractor fees	926
Contingency	4,814
Waste disposal indirect investment	1,129
Total fixed investment	33,356
Other Capital Investment	
Allowance for startup and modifications	2,888
Interest during construction	5,204
Land	377
Working capital	1,087
Total capital investment	42,912
$/kW	85.8

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-22. CASE 1, 500-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	275 tons	65	18
Total raw material cost			18
Conversion costs			
Operating labor and supervision			
Process	15,330 man-hr	15.00	230
Landfill	20,742 man-hr	21.00	436
Utilities			
Process water	7,549 kgal	0.14	1
Electricity	15,733,157 kWh	0.037	582
Diesel fuel	84,360 gal	1.60	135
Maintenance			
Labor and material			1,026
Analysis	300 man-hr	21.00	6
Total conversion costs			2,416
Total direct costs			2,434
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,019
Total first-year operating and maintenance costs			3,453
Levelized capital charges (14.7% of total capital investment)			6,308
Total first-year annual revenue requirements			9,761
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			6,512
Levelized capital charges (14.7% of total capital investment)			6,308
Total levelized annual revenue requirements			12,820

	M$	Mills/kWh
First-year annual revenue requirements	9.8	3.5
Levelized annual revenue requirements	12.8	4.7

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-23. CASE 1, 500-MW, TOTAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
NO_x removal areas	15,547
SO_2 removal areas	49,035
Particulate removal areas	17,467
Total process capital	82,049
Services, utilities, and miscellaneous	4,923
Total direct investment excluding waste disposal	86,972
Waste disposal	7,379
Total direct investment	94,351
Indirect Investment	
Engineering design and supervision	6,088
Architect and engineering contractor	1,740
Construction expense	13,915
Contractor fees	4,349
Contingency	15,856
Waste disposal indirect investment	2,492
Total fixed investment	138,791
Other Capital Investment	
Allowance for startup and modifications	11,405
Interest during construction	21,652
Royalties	463
Land	848
Working capital	5,172
Catalyst	14,678
Total capital investment	193,009
$/kW	386.0

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-24. CASE 1, 500-MW, TOTAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,640 tons	155	409
Catalyst	720 tons	23,558	16,962
Sodium hydroxide	39 tons	300	12
H_2SO_4 (100% equivalent)	295 tons	65	19
Limestone	143,058 tons	8.50	1,216
Total raw material cost			18,618
Conversion costs			
Operating labor and supervision			
Process	63,570 man-hr	15.00	954
Landfill	45,787 man-hr	21.00	962
Utilities			
Steam	432,123 MBtu	3.30	1,426
Process water	204,208 kgal	0.14	28
Electricity	81,443,366 kWh	0.037	3,013
Diesel fuel	186,222 gal	1.60	298
Maintenance			
Labor and material			5,964
Analysis	7,430 man-hr	21.00	156
Total conversion costs			12,801
Total direct costs			31,419
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,822
Total first-year operating and maintenance costs			36,241
Levelized capital charges (14.7% of total capital investment)			28,372
Total first-year annual revenue requirements			64,613
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			68,350
Levelized capital charges (14.7% of total capital investment)			28,372
Total levelized annual revenue requirements			96,722

	M$	Mills/kWh
First-year annual revenue requirements	64.6	23.5
Levelized annual revenue requirements	96.7	35.2

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-25. CASE 1, 1,000-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	1,997
Reactor	14,937
Flue gas handling	7,561
Air heater	1,606
Total process capital	26,101
Services, utilities, and miscellaneous	1,566
Total direct investment excluding waste disposal	27,667
Waste disposal	33
Total direct investment	27,700
Indirect Investment	
Engineering design and supervision	1,660
Architect and engineering contractor	277
Construction expense	3,873
Contractor fees	1,107
Contingency	6,917
Waste disposal indirect investment	10
Total fixed investment	41,544
Other Capital Investment	
Allowance for startup and modifications	4,150
Interest during construction	6,481
Royalties	902
Land	16
Working capital	1,170
Catalyst	23,444
Total capital investment	77,707
$/kW	77.7

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-26. CASE 1, 1,000-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	4,551 tons	155	705
Catalyst	1,150 tons	23,558	27,092
Sodium hydroxide	76 tons	300	23
H_2SO_4 (100% equivalent)	38 tons	65	2
Total raw material cost			27,822
Conversion costs			
Operating labor and supervision			
Process	6,570 man-hr	15.00	99
Landfill	180 man-hr	21.00	4
Utilities			
Steam	30,162 MBtu	3.30	100
Process water	5,092 kgal	0.14	1
Electricity	14,543,422 kWh	0.037	538
Diesel fuel	953 gal	1.60	2
Maintenance			
Labor and material			831
Analysis	2,628 man-hr	21.00	55
Total conversion costs			1,630
Total direct costs			29,452
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			593
Total first-year operating and maintenance costs			30,045
Levelized capital charges (14.7% of total capital investment)			11,423
Total first-year annual revenue requirements			41,468
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			56,665
Levelized capital charges (14.7% of total capital investment)			11,423
Total levelized annual revenue requirements			68,088

	M$	Mills/kWh
First-year annual revenue requirements	41.5	7.5
Levelized annual revenue requirements	68.1	12.4

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-27. CASE 1, 1,000-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	2,916
Feed preparation	6,042
Flue gas handling	21,830
SO_2 absorption	40,107
Oxidation	5,556
Reheat	7,153
Solids separation	4,717
Total process capital	88,321
Services, utilities, and miscellaneous	5,299
Total direct investment excluding waste disposal	93,620
Waste disposal	6,620
Total direct investment	100,240
Indirect Investment	
Engineering design and supervision	5,617
Architect and engineering contractor	936
Construction expense	13,107
Contractor fees	3,745
Contingency	11,703
Waste disposal indirect investment	2,076
Total fixed investment	137,424
Other Capital Investment	
Allowance for startup and modifications	10,298
Interest during construction	21,438
Land	708
Working capital	5,783
Total capital investment	175,651
$/kW	175.7

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-28. CASE 1, 1,000-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Limestone	276,930 tons	8.50	2,354
Total raw material cost			2,354
Conversion costs			
Operating labor and supervision			
Process	55,130 man-hr	15.00	827
Landfill	36,162 man-hr	21.00	759
Utilities			
Steam	803,333 MBtu	3.30	2,651
Process water	375,460 kgal	0.14	53
Electricity	109,941,621 kWh	0.037	4,068
Diesel fuel	190,989 gal	1.60	306
Maintenance			
Labor and material			6,752
Analysis	6,590 man-hr	21.00	138
Total conversion costs			15,554
Total direct costs			17,908
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			5,086
Total first-year operating and maintenance costs			22,994
Levelized capital charges (14.7% of total capital investment)			25,821
Total first-year annual revenue requirements			48,815
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			43,367
Levelized capital charges (14.7% of total capital investment)			25,821
Total levelized annual revenue requirements			69,188

	M$	Mills/kWh
First-year annual revenue requirements	48.8	8.9
Levelized annual revenue requirements	69.2	12.6

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-30. CASE 1, 1,000-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	533 tons	65	35
Total raw material cost			35
Conversion costs			
Operating labor and supervision			
Process	21,900 man-hr	15.00	329
Landfill	30,218 man-hr	21.00	635
Utilities			
Process water	14,621 kgal	0.14	2
Electricity	30,395,983 kWh	0.037	1,125
Diesel fuel	159,597 gal	1.60	255
Maintenance			
Labor and material			1,492
Analysis	400 man-hr	21.00	8
Total conversion costs			3,846
Total direct costs			3,881
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,478
Total first-year operating and maintenance costs			5,359
Levelized capital charges (14.7% of total capital investment)			10,768
Total first-year annual revenue requirements			16,127
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			10,107
Levelized capital charges (14.7% of total capital investment)			10,768
Total levelized annual revenue requirements			20,875

	M$	Mills/kWh
First-year annual revenue requirements	16.1	2.9
Levelized annual revenue requirements	20.9	3.8

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-31. CASE 1, 1,000-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	26,101
SO_2 removal areas	88,321
Particulate removal areas	31,261
Total process capital	145,683
Services, utilities, and miscellaneous	8,741
Total direct investment excluding waste disposal	154,424
Waste disposal	12,189
Total direct investment	166,613
Indirect Investment	
Engineering design and supervision	9,265
Architect and engineering contractor	1,544
Construction expense	21,619
Contractor fees	6,177
Contingency	26,904
Waste disposal indirect investment	3,821
Total fixed investment	235,943
Other Capital Investment	
Allowance for startup and modifications	19,418
Interest during construction	36,807
Royalties	902
Land	1,312
Working capital	8,782
Catalyst	23,444
Total capital investment	326,608
$/kW	326.6

Basis: 3.5% sulfur bituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, forced-oxidation limestone FGD and cold-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-32. CASE 1, 1,000-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	4,551 tons	155	705
Catalyst	1,150 tons	23,558	27,092
Sodium hydroxide	76 tons	300	23
H_2SO_4 (100% equivalent)	571 tons	65	37
Limestone	276,930 tons	8.50	2,354
Total raw material cost			30,211
Conversion costs			
Operating labor and supervision			
Process	83,600 man-hr	15.00	1,255
Landfill	66,560 man-hr	21.00	1,398
Utilities			
Steam	833,495 MBtu	3.30	2,751
Process water	395,173 kgal	0.14	56
Electricity	154,881,026 kWh	0.037	5,731
Diesel fuel	351,539 gal	1.60	563
Maintenance			
Labor and material			9,075
Analysis	9,618 man-hr	21.00	201
Total conversion costs			21,030
Total direct costs			51,241
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			7,157
Total first-year operating and maintenance costs			58,398
Levelized capital charges (14.7% of total capital investment)			48,012
Total first-year annual revenue requirements			106,410
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			110,139
Levelized capital charges (14.7% of total capital investment)			48,012
Total levelized annual revenue requirements			158,151

	M$	Mills/kWh
First-year annual revenue requirements	106.4	19.3
Levelized annual revenue requirements	158.2	28.8

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-33. CASE 2, 200-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	838
Reactor	3,743
Flue gas handling	2,876
Air heater	713
Total process capital	8,170
Services, utilities, and miscellaneous	490
Total direct investment excluding waste disposal	8,660
Waste disposal	19
Total direct investment	8,679
Indirect Investment	
Engineering design and supervision	693
Architect and engineering contractor	260
Construction expense	1,559
Contractor fees	520
Contingency	2,338
Waste disposal indirect investment	7
Total fixed investment	14,056
Other Capital Investment	
Allowance for startup and modifications	1,403
Interest during construction	2,193
Royalties	231
Land	8
Working capital	406
Catalyst	5,912
Total capital investment	24,209
$/kW	121.0

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-34. CASE 2, 200-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	1,059 tons	155	164
Catalyst	290 tons	23,558	6,832
Sodium hydroxide	21 tons	300	6
H_2SO_4 (100% equivalent)	10 tons	65	1
Total raw material cost			7,003
Conversion costs			
Operating labor and supervision			
Process	2,190 man-hr	15.00	33
Landfill	256 man-hr	21.00	5
Utilities			
Steam	8,914 MBtu	3.30	29
Process water	1,380 kgal	0.14	0
Electricity	5,511,653 kWh	0.037	204
Diesel fuel	425 gal	1.60	1
Maintenance			
Labor and material			434
Analysis	1,752 man-hr	21.00	37
Total conversion costs			743
Total direct costs			7,746
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			305
Total first-year operating and maintenance costs			8,051
Levelized capital charges (14.7% of total capital investment)			3,559
Total first-year annual revenue requirements			11,610
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			15,184
Levelized capital charges (14.7% of total capital investment)			3,559
Total levelized annual revenue requirements			18,743

	M$	Mills/kWh
First-year annual revenue requirements	11.6	10.6
Levelized annual revenue requirements	18.7	17.0

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-35. CASE 2, 200-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	694
Feed preparation	790
Flue gas handling	3,956
SO_2 absorption	7,476
Lime particulate recycle	1,153
Total process capital	14,069
Services, utilities, and miscellaneous	844
Total direct investment excluding waste disposal	14,913
Waste disposal	299
Total direct investment	15,212
Indirect Investment	
Engineering design and supervision	1,193
Architect and engineering contractor	447
Construction expense	2,684
Contractor fees	895
Contingency	4,026
Waste disposal indirect investment	107
Total fixed investment	24,564
Other Capital Investment	
Allowance for startup and modifications	2,416
Interest during construction	3,832
Land	41
Working capital	806
Total capital investment	31,659
$/kW	158.3

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-36. CASE 2, 200-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Lime	3,845 tons	75	288
Total raw material cost			288
Conversion costs			
Operating labor and supervision			
Process	13,140 man-hr	15.00	197
Landfill	3,970 man-hr	21.00	83
Utilities			
Process water	47,065 kgal	0.14	7
Electricity	8,576,199 kWh	0.037	317
Diesel fuel	6,577 gal	1.60	11
Maintenance			
Labor and material			1,053
Analysis	4,191 man-hr	21.00	88
Total conversion costs			1,756
Total direct costs			2,044
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			853
Total first-year operating and maintenance costs			2,897
Levelized capital charges (14.7% of total capital investment)			4,654
Total first-year annual revenue requirements			7,551
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			5,464
Levelized capital charges (14.7% of total capital investment)			4,654
Total levelized annual revenue requirements			10,118

	M$	Mills/kWh
First-year annual revenue requirements	7.6	6.9
Levelized annual revenue requirements	10.1	9.2

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-37. CASE 2, 200-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Particulate removal and storage	7,078
Particulate transfer	3,704
Flue gas handling	2,185
Total process capital	12,967
Services, utilities, and miscellaneous	778
Total direct investment excluding waste disposal	13,745
Waste disposal	1,563
Total direct investment	15,308
Indirect Investment	
Engineering design and supervision	1,100
Architect and engineering contractor	412
Construction expense	2,474
Contractor fees	825
Contingency	3,711
Waste disposal indirect investment	561
Total fixed investment	24,391
Other Capital Investment	
Allowance for startup and modifications	2,227
Interest during construction	3,805
Land	181
Working capital	839
Total capital investment	31,443
$/kW	157.2

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-38. CASE 2, 200-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	224 tons	65	15
Total raw material cost			15
Conversion costs			
Operating labor and supervision			
Process	10,950 man-hr	15.00	164
Landfill	20,734 man-hr	21.00	435
Utilities			
Process water	950 kgal	0.14	0
Electricity	10,676,617 kWh	0.037	395
Diesel fuel	34,348 gal	1.60	55
Maintenance			
Labor and material			1,009
Analysis	200 man-hr	21.00	4
Total conversion costs			2,062
Total direct costs			2,077
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			967
Total first-year operating and maintenance costs			3,044
Levelized capital charges (14.7% of total capital investment)			4,622
Total first-year annual revenue requirements			7,666
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			5,741
Levelized capital charges (14.7% of total capital investment)			4,622
Total levelized annual revenue requirements			10,363

	M$	Mills/kWh
First-year annual revenue requirements	7.7	7.0
Levelized annual revenue requirements	10.4	9.4

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-39. CASE 2, 200-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	8,170
SO_2 removal areas	14,069
Particulate removal areas	12,967
Total process capital	35,206
Services, utilities, and miscellaneous	2,112
Total direct investment excluding waste disposal	37,318
Waste disposal	1,881
Total direct investment	39,199
Indirect Investment	
Engineering design and supervision	2,986
Architect and engineering contractor	1,119
Construction expense	6,717
Contractor fees	2,240
Contingency	10,075
Waste disposal indirect investment	675
Total fixed investment	63,011
Other Capital Investment	
Allowance for startup and modifications	6,046
Interest during construction	9,830
Royalties	231
Land	230
Working capital	2,051
Catalyst	5,912
Total capital investment	87,311
$/kW	436.6

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-40. CASE 2, 200-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	1,059 tons	155	164
Catalyst	290 tons	23,558	6,832
Sodium hydroxide	21 tons	300	6
H_2SO_4 (100% equivalent)	235 tons	65	16
Lime	3,845 tons	75	288
Total raw material cost			7,306
Conversion costs			
Operating labor and supervision			
Process	26,280 man-hr	15.00	394
Landfill	24,960 man-hr	21.00	523
Utilities			
Steam	8,914 MBtu	3.30	29
Process water	49,395 kgal	0.14	7
Electricity	24,764,469 kWh	0.037	916
Diesel fuel	41,350 gal	1.60	67
Maintenance			
Labor and material			2,496
Analysis	6,143 man-hr	21.00	129
Total conversion costs			4,561
Total direct costs			11,867
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,125
Total first-year operating and maintenance costs			13,992
Levelized capital charges (14.7% of total capital investment)			12,835
Total first-year annual revenue requirements			26,827
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			26,389
Levelized capital charges (14.7% of total capital investment)			12,835
Total levelized annual revenue requirements			39,224

	M$	Mills/kWh
First-year annual revenue requirements	26.8	24.4
Levelized annual revenue requirements	39.2	35.7

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-41. CASE 2, 500-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	1,328
Reactor	9,278
Flue gas handling	4,543
Air heater	1,220
Total process capital	16,369
Services, utilities, and miscellaneous	982
Total direct investment excluding waste disposal	17,351
Waste disposal	34
Total direct investment	17,385
Indirect Investment	
Engineering design and supervision	1,215
Architect and engineering contractor	347
Construction expense	2,776
Contractor fees	868
Contingency	4,511
Waste disposal indirect investment	12
Total fixed investment	27,114
Other Capital Investment	
Allowance for startup and modifications	2,707
Interest during construction	4,230
Royalties	563
Land	15
Working capital	783
Catalyst	14,678
Total capital investment	50,090
$/kW	100.2

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-42. CASE 2, 500-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,167 tons	155	336
Catalyst	720 tons	23,558	16,962
Sodium hydroxide	50 tons	300	15
H_2SO_4 (100% equivalent)	25 tons	65	2
Total raw material cost			17,315
Conversion costs			
Operating labor and supervision			
Process	4,380 man-hr	15.00	66
Landfill	257 man-hr	21.00	5
Utilities			
Steam	19,819 MBtu	3.30	65
Process water	3,382 kgal	0.14	0
Electricity	13,305,600 kWh	0.037	492
Diesel fuel	738 gal	1.60	1
Maintenance			
Labor and material			695
Analysis	2,190 man-hr	21.00	46
Total conversion costs			1,370
Total direct costs			18,685
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			487
Total first-year operating and maintenance costs			19,172
Levelized capital charges (14.7% of total capital investment)			7,363
Total first-year annual revenue requirements			26,535
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			36,158
Levelized capital charges (14.7% of total capital investment)			7,363
Total levelized annual revenue requirements			43,521

	M$	Mills/kWh
First-year annual revenue requirements	26.5	9.6
Levelized annual revenue requirements	43.5	15.8

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-43. CASE 2, 500-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	1,132
Feed preparation	1,258
Flue gas handling	7,374
SO_2 absorption	12,992
Lime particulate recycle	2,140
Total process capital	24,896
Services, utilities, and miscellaneous	1,494
Total direct investment excluding waste disposal	26,390
Waste disposal	527
Total direct investment	26,917
Indirect Investment	
Engineering design and supervision	1,847
Architect and engineering contractor	528
Construction expense	4,222
Contractor fees	1,320
Contingency	6,861
Waste disposal indirect investment	181
Total fixed investment	41,876
Other Capital Investment	
Allowance for startup and modifications	4,117
Interest during construction	6,533
Land	75
Working capital	1,375
Total capital investment	53,976
$/kW	108.0

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-44. CASE 2, 500-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Lime	9,446 tons	75	708
Total raw material cost			708
Conversion costs			
Operating labor and supervision			
Process	17,520 man-hr	15.00	263
Landfill	3,976 man-hr	21.00	83
Utilities			
Process water	115,303 kgal	0.14	16
Electricity	21,088,330 kWh	0.037	780
Diesel fuel	11,380 gal	1.60	18
Maintenance			
Labor and material			1,599
Analysis	4,191 man-hr	21.00	88
Total conversion costs			2,847
Total direct costs			3,555
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,220
Total first-year operating and maintenance costs			4,775
Levelized capital charges (14.7% of total capital investment)			7,934
Total first-year annual revenue requirements			12,709
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			9,006
Levelized capital charges (14.7% of total capital investment)			7,934
Total levelized annual revenue requirements			16,940

	M$	Mills/kWh
First-year annual revenue requirements	12.7	4.6
Levelized annual revenue requirements	16.9	6.2

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-45. CASE 2, 500-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Particulate removal and storage	15,446
Particulate transfer	6,779
Flue gas handling	4,961
Total process capital	27,186
Services, utilities, and miscellaneous	1,631
Total direct investment excluding waste disposal	28,817
Waste disposal	2,749
Total direct investment	31,566
Indirect Investment	
Engineering design and supervision	2,017
Architect and engineering contractor	576
Construction expense	4,611
Contractor fees	1,441
Contingency	7,492
Waste disposal indirect investment	944
Total fixed investment	48,647
Other Capital Investment	
Allowance for startup and modifications	4,495
Interest during construction	7,589
Land	326
Working capital	1,592
Total capital investment	62,649
$/kW	125.3

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-46. CASE 2, 500-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	550 tons	65	36
Total raw material cost			36
Conversion costs			
Operating labor and supervision			
Process	19,710 man-hr	15.00	296
Landfill	20,726 man-hr	21.00	435
Utilities			
Process water	2,330 kgal	0.14	0
Electricity	26,111,655 kWh	0.037	966
Diesel fuel	59,320 gal	1.60	95
Maintenance			
Labor and material			1,811
Analysis	300 man-hr	21.00	6
Total conversion costs			3,609
Total direct costs			3,645
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,529
Total first-year operating and maintenance costs			5,174
Levelized capital charges (14.7% of total capital investment)			9,209
Total first-year annual revenue requirements			14,383
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			9,758
Levelized capital charges (14.7% of total capital investment)			9,209
Total levelized annual revenue requirements			18,967

	M$	Mills/kWh
First-year annual revenue requirements	14.4	5.2
Levelized annual revenue requirements	19.0	6.9

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-47. CASE 2, 500-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	16,369
SO_2 removal areas	24,896
Particulate removal areas	27,186
Total process capital	68,451
Services, utilities, and miscellaneous	4,107
Total direct investment excluding waste disposal	72,558
Waste disposal	3,310
Total direct investment	75,868
Indirect Investment	
Engineering design and supervision	5,079
Architect and engineering contractor	1,451
Construction expense	11,609
Contractor fees	3,629
Contingency	18,864
Waste disposal indirect investment	1,137
Total fixed investment	117,637
Other Capital Investment	
Allowance for startup and modifications	11,319
Interest during construction	18,352
Royalties	563
Land	416
Working capital	3,750
Catalyst	14,678
Total capital investment	166,715
$/kW	333.4

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-48. CASE 2, 500-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,167 tons	155	336
Catalyst	720 tons	23,558	16,962
Sodium hydroxide	50 tons	300	15
H_2SO_4 (100% equivalent)	575 tons	65	38
Lime	9,446 tons	75	708
Total raw material cost			18,059
Conversion costs			
Operating labor and supervision			
Process	41,610 man-hr	15.00	625
Landfill	24,959 man-hr	21.00	523
Utilities			
Steam	19,819 MBtu	3.30	65
Process water	121,015 kgal	0.14	16
Electricity	60,505,585 kWh	0.037	2,238
Diesel fuel	71,438 gal	1.60	114
Maintenance			
Labor and material			4,105
Analysis	6,681 man-hr	21.00	140
Total conversion costs			7,826
Total direct costs			25,885
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,236
Total first-year operating and maintenance costs			29,121
Levelized capital charges (14.7% of total capital investment)			24,506
Total first-year annual revenue requirements			53,627
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			54,922
Levelized capital charges (14.7% of total capital investment)			24,506
Total levelized annual revenue requirements			79,428

	M$	Mills/kWh
First-year annual revenue requirements	53.6	19.5
Levelized annual revenue requirements	79.4	28.9

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-49. CASE 2, 500-MW, NO_x REMOVAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	1,431
Reactor	10,654
Flue gas handling	4,545
Air heater	1,221
Total process capital	17,851
Services, utilities, and miscellaneous	1,071
Total direct investment excluding waste disposal	18,922
Waste disposal	39
Total direct investment	18,961
Indirect Investment	
Engineering design and supervision	1,325
Architect and engineering contractor	378
Construction expense	3,028
Contractor fees	946
Contingency	4,920
Waste disposal indirect investment	13
Total fixed investment	29,571
Other Capital Investment	
Allowance for startup and modifications	2,952
Interest during construction	4,613
Royalties	563
Land	22
Working capital	848
Catalyst	16,920
Total capital investment	55,489
$/kW	111.0

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-50. CASE 2, 500-MW, NO$_x$ REMOVAL ANNUAL REVENUE REQUIREMENTS, 90% NO$_x$ REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,432 tons	155	377
Catalyst	830 tons	23,558	19,553
Sodium hydroxide	50 tons	300	15
H_2SO_4 (100% equivalent)	25 tons	65	2
Total raw material cost			19,947
Conversion costs			
Operating labor and supervision			
Process	4,380 man-hr	15.00	66
Landfill	296 man-hr	21.00	6
Utilities			
Steam	21,083 MBtu	3.30	70
Process water	3,387 kgal	0.14	0
Electricity	13,438,932 kWh	0.037	497
Diesel fuel	848 gal	1.60	1
Maintenance			
Labor and material			758
Analysis	2,190 man-hr	21.00	46
Total conversion costs			1,444
Total direct costs			21,391
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			526
Total first-year operating and maintenance costs			21,917
Levelized capital charges (14.7% of total capital investment)			8,157
Total first-year annual revenue requirements			30,074
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			41,335
Levelized capital charges (14.7% of total capital investment)			8,157
Total levelized annual revenue requirements			49,492

	M$	Mills/kWh
First-year annual revenue requirements	30.1	10.9
Levelized annual revenue requirements	49.5	18.0

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-51. CASE 2, 500-MW, SO_2 REMOVAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
Materials handling	1,132
Feed preparation	1,258
Flue gas handling	7,377
SO_2 absorption	12,997
Lime particulate recycle	2,142
Total process capital	24,906
Services, utilities, and miscellaneous	1,494
Total direct investment excluding waste disposal	26,400
Waste disposal	527
Total direct investment	26,927
Indirect Investment	
Engineering design and supervision	1,848
Architect and engineering contractor	528
Construction expense	4,224
Contractor fees	1,320
Contingency	6,864
Waste disposal indirect investment	181
Total fixed investment	41,892
Other Capital Investment	
Allowance for startup and modifications	4,118
Interest during construction	6,535
Land	75
Working capital	1,375
Total capital investment	53,995
$/kW	108.0

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-52. CASE 2, 500-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Lime	9,446 tons	75	708
Total raw material cost			708
Conversion costs			
Operating labor and supervision			
Process	17,520 man-hr	15.00	263
Landfill	3,976 man-hr	21.00	83
Utilities			
Process water	115,398 kgal	0.14	16
Electricity	21,104,983 kWh	0.037	781
Diesel fuel	11,380 gal	1.60	18
Maintenance			
Labor and material			1,600
Analysis	4,191 man-hr	21.00	88
Total conversion costs			2,849
Total direct costs			3,557
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,220
Total first-year operating and maintenance costs			4,777
Levelized capital charges (14.7% of total capital investment)			7,937
Total first-year annual revenue requirements			12,714
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			9,009
Levelized capital charges (14.7% of total capital investment)			7,937
Total levelized annual revenue requirements			16,946

	M$	Mills/kWh
First-year annual revenue requirements	12.7	4.6
Levelized annual revenue requirements	16.9	6.2

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-53. CASE 2, 500-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
Particulate removal and storage	15,457
Particulate transfer	6,779
Flue gas handling	4,964
Total process capital	27,200
Services, utilities, and miscellaneous	1,632
Total direct investment excluding waste disposal	28,832
Waste disposal	2,749
Total direct investment	31,581
Indirect Investment	
Engineering design and supervision	2,018
Architect and engineering contractor	577
Construction expense	4,613
Contractor fees	1,442
Contingency	7,496
Waste disposal indirect investment	944
Total fixed investment	48,671
Other Capital Investment	
Allowance for startup and modifications	4,498
Interest during construction	7,593
Land	326
Working capital	1,593
Total capital investment	62,681
$/kW	125.4

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-54. CASE 2, 500-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	550 tons	65	36
Total raw material cost			36
Conversion costs			
Operating labor and supervision			
Process	19,710 man-hr	15.00	296
Landfill	20,726 man-hr	21.00	435
Utilities			
Process water	2,330 kgal	0.14	0
Electricity	26,134,704 kWh	0.037	967
Diesel fuel	59,320 gal	1.60	95
Maintenance			
Labor and material			1,812
Analysis	300 man-hr	21.00	6
Total conversion costs			3,611
Total direct costs			3,647
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,529
Total first-year operating and maintenance costs			5,176
Levelized capital charges (14.7% of total capital investment)			9,214
Total first-year annual revenue requirements			14,390
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			9,762
Levelized capital charges (14.7% of total capital investment)			9,214
Total levelized annual revenue requirements			18,976

	M$	Mills/kWh
First-year annual revenue requirements	14.4	5.2
Levelized annual revenue requirements	19.0	6.9

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-55. CASE 2, 500-MW, TOTAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
NO_x removal areas	17,851
SO_2 removal areas	24,906
Particulate removal areas	27,200
Total process capital	69,957
Services, utilities, and miscellaneous	4,197
Total direct investment excluding waste disposal	74,154
Waste disposal	3,315
Total direct investment	77,469
Indirect Investment	
Engineering design and supervision	5,191
Architect and engineering contractor	1,483
Construction expense	11,865
Contractor fees	3,708
Contingency	19,280
Waste disposal indirect investment	1,138
Total fixed investment	120,134
Other Capital Investment	
Allowance for startup and modifications	11,568
Interest during construction	18,741
Royalties	563
Land	423
Working capital	3,816
Catalyst	16,920
Total capital investment	172,165
$/kW	344.3

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-56. CASE 2, 500-MW, TOTAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,432 tons	155	377
Catalyst	830 tons	23,558	19,553
Sodium hydroxide	50 tons	300	15
H_2SO_4 (100% equivalent)	575 tons	65	38
Lime	9,446 tons	75	708
Total raw material cost			20,691
Conversion costs			
Operating labor and supervision			
Process	41,610 man-hr	15.00	625
Landfill	24,998 man-hr	21.00	524
Utilities			
Steam	21,083 MBtu	3.30	70
Process water	121,115 kgal	0.14	16
Electricity	60,678,619 kWh	0.037	2,245
Diesel fuel	71,548 gal	1.60	114
Maintenance			
Labor and material			4,170
Analysis	6,681 man-hr	21.00	140
Total conversion costs			7,904
Total direct costs			28,595
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,275
Total first-year operating and maintenance costs			31,870
Levelized capital charges (14.7% of total capital investment)			25,308
Total first-year annual revenue requirements			57,178
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			60,106
Levelized capital charges (14.7% of total capital investment)			25,308
Total levelized annual revenue requirements			85,414

	M$	Mills/kWh
First-year annual revenue requirements	57.2	20.8
Levelized annual revenue requirements	85.4	31.1

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-57. CASE 2, 1,000-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	2,290
Reactor	17,988
Flue gas handling	8,960
Air heater	2,400
Total process capital	31,638
Services, utilities, and miscellaneous	1,898
Total direct investment excluding waste disposal	33,536
Waste disposal	54
Total direct investment	33,590
Indirect Investment	
Engineering design and supervision	2,012
Architect and engineering contractor	335
Construction expense	4,695
Contractor fees	1,341
Contingency	8,384
Waste disposal indirect investment	18
Total fixed investment	50,375
Other Capital Investment	
Allowance for startup and modifications	5,030
Interest during construction	7,859
Royalties	1,091
Land	25
Working capital	1,449
Catalyst	28,948
Total capital investment	94,777
$/kW	94.8

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-58. CASE 2, 1,000-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	5,046 tons	155	782
Catalyst	1,420 tons	23,558	33,452
Sodium hydroxide	98 tons	300	29
H_2SO_4 (100% equivalent)	49 tons	65	3
Total raw material cost			34,266
Conversion costs			
Operating labor and supervision			
Process	6,570 man-hr	15.00	99
Landfill	350 man-hr	21.00	7
Utilities			
Steam	42,480 MBtu	3.30	140
Process water	6,579 kgal	0.14	1
Electricity	26,269,165 kWh	0.037	972
Diesel fuel	1,254 gal	1.60	2
Maintenance			
Labor and material			1,008
Analysis	2,628 man-hr	21.00	55
Total conversion costs			2,284
Total direct costs			36,550
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			701
Total first-year operating and maintenance costs			37,251
Levelized capital charges (14.7% of total capital investment)			13,932
Total first-year annual revenue requirements			51,183
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			70,255
Levelized capital charges (14.7% of total capital investment)			13,932
Total levelized annual revenue requirements			84,187

	M$	Mills/kWh
First-year annual revenue requirements	51.2	9.3
Levelized annual revenue requirements	84.2	15.3

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-59. CASE 2, 1,000-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	1,630
Feed preparation	1,780
Flue gas handling	14,524
SO_2 absorption	25,598
Lime particulate recycle	3,342
Total process capital	46,874
Services, utilities, and miscellaneous	2,812
Total direct investment excluding waste disposal	49,686
Waste disposal	831
Total direct investment	50,517
Indirect Investment	
Engineering design and supervision	2,981
Architect and engineering contractor	497
Construction expense	6,956
Contractor fees	1,987
Contingency	12,421
Waste disposal indirect investment	269
Total fixed investment	75,628
Other Capital Investment	
Allowance for startup and modifications	7,453
Interest during construction	11,798
Land	119
Working capital	2,442
Total capital investment	97,440
$/kW	97.4

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-60. CASE 2, 1,000-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Lime	18,489 tons	75	1,387
Total raw material cost			1,387
Conversion costs			
Operating labor and supervision			
Process	21,900 man-hr	15.00	329
Landfill	5,332 man-hr	21.00	112
Utilities			
Process water	224,300 kgal	0.14	31
Electricity	41,005,970 kWh	0.037	1,517
Diesel fuel	19,078 gal	1.60	31
Maintenance			
Labor and material			2,509
Analysis	5,590 man-hr	21.00	117
Total conversion costs			4,646
Total direct costs			6,033
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,840
Total first-year operating and maintenance costs			7,873
Levelized capital charges (14.7% of total capital investment)			14,324
Total first-year annual revenue requirements			22,197
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			14,848
Levelized capital charges (14.7% of total capital investment)			14,324
Total levelized annual revenue requirements			29,172

	M$	Mills/kWh
First-year annual revenue requirements	22.2	4.0
Levelized annual revenue requirements	29.2	5.3

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-61. CASE 2, 1,000-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Particulate removal and storage	30,145
Particulate transfer	10,609
Flue gas handling	9,752
Total process capital	50,506
Services, utilities, and miscellaneous	3,030
Total direct investment excluding waste disposal	53,536
Waste disposal	4,305
Total direct investment	57,841
Indirect Investment	
Engineering design and supervision	3,212
Architect and engineering contractor	535
Construction expense	7,495
Contractor fees	2,141
Contingency	13,384
Waste disposal indirect investment	1,395
Total fixed investment	86,003
Other Capital Investment	
Allowance for startup and modifications	8,030
Interest during construction	13,416
Land	508
Working capital	2,760
Total capital investment	110,717
$/kW	110.7

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-62. CASE 2, 1,000-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	1,069 tons	65	69
Total raw material cost			69
Conversion costs			
Operating labor and supervision			
Process	28,470 man-hr	15.00	427
Landfill	27,597 man-hr	21.00	580
Utilities			
Process water	4,526 kgal	0.14	1
Electricity	50,762,221 kWh	0.037	1,878
Diesel fuel	98,740 gal	1.60	158
Maintenance			
Labor and material			2,806
Analysis	400 man-hr	21.00	8
Total conversion costs			5,858
Total direct costs			5,927
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,293
Total first-year operating and maintenance costs			8,220
Levelized capital charges (14.7% of total capital investment)			16,275
Total first-year annual revenue requirements			24,495
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			15,503
Levelized capital charges (14.7% of total capital investment)			16,275
Total levelized annual revenue requirements			31,778

	M$	Mills/kWh
First-year annual revenue requirements	24.5	4.5
Levelized annual revenue requirements	31.8	5.8

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-63. CASE 2, 1,000-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	31,638
SO_2 removal areas	46,874
Particulate removal areas	50,506
Total process capital	129,018
Services, utilities, and miscellaneous	7,740
Total direct investment excluding waste disposal	136,758
Waste disposal	5,190
Total direct investment	141,948
Indirect Investment	
Engineering design and supervision	8,205
Architect and engineering contractor	1,367
Construction expense	19,146
Contractor fees	5,469
Contingency	34,189
Waste disposal indirect investment	1,682
Total fixed investment	212,006
Other Capital Investment	
Allowance for startup and modifications	20,513
Interest during construction	33,073
Royalties	1,091
Land	652
Working capital	6,651
Catalyst	28,948
Total capital investment	302,934
$/kW	302.9

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, lime spray dryer FGD and baghouse to meet 1979 NSPS, mid-1982 costs.

TABLE A-64. CASE 2, 1,000-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	5,046 tons	155	782
Catalyst	1,420 tons	23,558	33,452
Sodium hydroxide	98 tons	300	29
H_2SO_4 (100% equivalent)	1,118 tons	65	72
Lime	18,489 tons	75	1,387
Total raw material cost			35,722
Conversion costs			
Operating labor and supervision			
Process	56,940 man-hr	15.00	855
Landfill	33,279 man-hr	21.00	699
Utilities			
Steam	42,480 MBtu	3.30	140
Process water	235,405 kgal	0.14	33
Electricity	118,037,356 kWh	0.037	4,367
Diesel fuel	119,072 gal	1.60	191
Maintenance			
Labor and material			6,323
Analysis	8,618 man-hr	21.00	180
Total conversion costs			12,788
Total direct costs			48,510
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,834
Total first-year operating and maintenance costs			53,344
Levelized capital charges (14.7% of total capital investment)			44,531
Total first-year annual revenue requirements			97,875
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			100,606
Levelized capital charges (14.7% of total capital investment)			44,531
Total levelized annual revenue requirements			145,137

	M$	Mills/kWh
First-year annual revenue requirements	97.9	17.8
Levelized annual revenue requirements	145.1	26.4

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-65. CASE 3, 200-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	830
Reactor	3,656
Flue gas handling	3,420
Air heater	503
Total process capital	8,409
Services, utilities, and miscellaneous	505
Total direct investment excluding waste disposal	8,914
Waste disposal	17
Total direct investment	8,931
Indirect Investment	
Engineering design and supervision	713
Architect and engineering contractor	267
Construction expense	1,605
Contractor fees	535
Contingency	2,407
Waste disposal indirect investment	6
Total fixed investment	14,464
Other Capital Investment	
Allowance for startup and modifications	1,444
Interest during construction	2,256
Royalties	231
Land	8
Working capital	411
Catalyst	5,504
Total capital investment	24,318
$/kW	121.6

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-66. CASE 3, 200-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	1,059 tons	155	164
Catalyst	270 tons	23,558	6,361
Sodium hydroxide	49 tons	300	15
H_2SO_4 (100% equivalent)	24 tons	65	2
Total raw material cost			6,542
Conversion costs			
Operating labor and supervision			
Process	2,190 man-hr	15.00	33
Landfill	214 man-hr	21.00	4
Utilities			
Steam	8,658 MBtu	3.30	29
Process water	3,252 kgal	0.14	0
Electricity	4,396,004 kWh	0.037	163
Diesel fuel	356 gal	1.60	1
Maintenance			
Labor and material			446
Analysis	1,752 man-hr	21.00	37
Total conversion costs			713
Total direct costs			7,255
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			312
Total first-year operating and maintenance costs			7,567
Levelized capital charges (14.7% of total capital investment)			3,575
Total first-year annual revenue requirements			11,142
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			14,271
Levelized capital charges (14.7% of total capital investment)			3,575
Total levelized annual revenue requirements			17,846

	M$	Mills/kWh
First-year annual revenue requirements	11.1	10.1
Levelized annual revenue requirements	17.8	16.2

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-67. CASE 3, 200-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	987
Feed preparation	1,612
Flue gas handling	6,037
SO_2 absorption	9,833
Solids separation	1,739
Total process capital	20,208
Services, utilities, and miscellaneous	1,212
Total direct investment excluding waste disposal	21,420
Waste disposal	470
Total direct investment	21,890
Indirect Investment	
Engineering design and supervision	1,714
Architect and engineering contractor	643
Construction expense	3,856
Contractor fees	1,285
Contingency	2,892
Waste disposal indirect investment	170
Total fixed investment	32,450
Other Capital Investment	
Allowance for startup and modifications	2,545
Interest during construction	5,062
Land	67
Working capital	1,264
Total capital investment	41,388
$/kW	206.9

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-68. CASE 3, 200-MW, SO$_2$ REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Limestone	8,903 tons	8.50	76
Total raw material cost			76
Conversion costs			
Operating labor and supervision			
Process	30,660 man-hr	15.00	460
Landfill	5,935 man-hr	21.00	125
Utilities			
Process water	55,573 kgal	0.14	8
Electricity	16,924,291 kWh	0.037	626
Diesel fuel	9,886 gal	1.60	16
Maintenance			
Labor and material			1,942
Analysis	3,300 man-hr	21.00	69
Total conversion costs			3,246
Total direct costs			3,322
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,558
Total first-year operating and maintenance costs			4,880
Levelized capital charges (14.7% of total capital investment)			6,084
Total first-year annual revenue requirements			10,964
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			9,204
Levelized capital charges (14.7% of total capital investment)			6,084
Total levelized annual revenue requirements			15,288

	M$	Mills/kWh
First-year annual revenue requirements	11.0	10.0
Levelized annual revenue requirements	15.3	13.9

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-69. CASE 3, 200-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Particulate removal and storage	6,035
Particulate transfer	2,619
Flue gas handling	2,732
Total process capital	11,386
Services, utilities, and miscellaneous	683
Total direct investment excluding waste disposal	12,069
Waste disposal	1,490
Total direct investment	13,559
Indirect Investment	
Engineering design and supervision	966
Architect and engineering contractor	362
Construction expense	2,172
Contractor fees	724
Contingency	3,259
Waste disposal indirect investment	537
Total fixed investment	21,579
Other Capital Investment	
Allowance for startup and modifications	1,955
Interest during construction	3,366
Land	173
Working capital	718
Total capital investment	27,791
$/kW	139.0

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-70. CASE 3, 200-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	224 tons	65	15
Total raw material cost			15
Conversion costs			
Operating labor and supervision			
Process	6,570 man-hr	15.00	99
Landfill	18,811 man-hr	21.00	395
Utilities			
Process water	950 kgal	0.14	0
Electricity	10,969,496 kWh	0.037	406
Diesel fuel	31,335 gal	1.60	50
Maintenance			
Labor and material			769
Analysis	200 man-hr	21.00	4
Total conversion costs			1,723
Total direct costs			1,738
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			760
Total first-year operating and maintenance costs			2,498
Levelized capital charges (14.7% of total capital investment)			4,085
Total first-year annual revenue requirements			6,583
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			4,711
Levelized capital charges (14.7% of total capital investment)			4,085
Total levelized annual revenue requirements			8,796

	M$	Mills/kWh
First-year annual revenue requirements	6.6	6.0
Levelized annual revenue requirements	8.8	8.0

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-71. CASE 3, 200-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	8,409
SO_2 removal areas	20,208
Particulate removal areas	11,386
Total process capital	40,003
Services, utilities, and miscellaneous	2,400
Total direct investment excluding waste disposal	42,403
Waste disposal	1,977
Total direct investment	44,380
Indirect Investment	
Engineering design and supervision	3,393
Architect and engineering contractor	1,272
Construction expense	7,633
Contractor fees	2,544
Contingency	8,558
Waste disposal indirect investment	713
Total fixed investment	68,493
Other Capital Investment	
Allowance for startup and modifications	5,944
Interest during construction	10,684
Royalties	231
Land	248
Working capital	2,393
Catalyst	5,504
Total capital investment	93,497
$/kW	467.5

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-72. CASE 3, 200-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	1,059 tons	155	164
Catalyst	270 tons	23,558	6,361
Sodium hydroxide	49 tons	300	15
H_2SO_4 (100% equivalent)	249 tons	65	17
Limestone	8,903 tons	8.50	76
Total raw material cost			6,633
Conversion costs			
Operating labor and supervision			
Process	39,420 man-hr	15.00	592
Landfill	24,960 man-hr	21.00	524
Utilities			
Steam	8,658 MBtu	3.30	29
Process water	59,775 kgal	0.14	8
Electricity	32,289,791 kWh	0.037	1,195
Diesel fuel	41,577 gal	1.60	67
Maintenance			
Labor and material			3,157
Analysis	5,252 man-hr	21.00	110
Total conversion costs			5,682
Total direct costs			12,315
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,630
Total first-year operating and maintenance costs			14,945
Levelized capital charges (14.7% of total capital investment)			13,744
Total first-year annual revenue requirements			28,689
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			28,186
Levelized capital charges (14.7% of total capital investment)			13,744
Total levelized annual revenue requirements			41,930

	M$	Mills/kWh
First-year annual revenue requirements	28.7	26.1
Levelized annual revenue requirements	41.9	38.1

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-73. CASE 3, 500-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	1,297
Reactor	8,453
Flue gas handling	5,386
Air heater	861
Total process capital	15,997
Services, utilities, and miscellaneous	960
Total direct investment excluding waste disposal	16,957
Waste disposal	30
Total direct investment	16,987
Indirect Investment	
Engineering design and supervision	1,187
Architect and engineering contractor	339
Construction expense	2,713
Contractor fees	848
Contingency	4,409
Waste disposal indirect investment	10
Total fixed investment	26,493
Other Capital Investment	
Allowance for startup and modifications	2,645
Interest during construction	4,133
Royalties	563
Land	15
Working capital	757
Catalyst	13,455
Total capital investment	48,061
$/kW	96.1

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-74. CASE 3, 500-MW, NO$_x$ REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,165 tons	155	336
Catalyst	660 tons	23,558	15,549
Sodium hydroxide	119 tons	300	36
H_2SO_4 (100% equivalent)	60 tons	65	4
Total raw material cost			15,925
Conversion costs			
Operating labor and supervision			
Process	4,380 man-hr	15.00	66
Landfill	214 man-hr	21.00	4
Utilities			
Steam	19,190 MBtu	3.30	63
Process water	7,969 kgal	0.14	1
Electricity	10,572,460 kWh	0.037	391
Diesel fuel	620 gal	1.60	1
Maintenance			
Labor and material			679
Analysis	2,190 man-hr	21.00	46
Total conversion costs			1,251
Total direct costs			17,176
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			477
Total first-year operating and maintenance costs			17,653
Levelized capital charges (14.7% of total capital investment)			7,065
Total first-year annual revenue requirements			24,718
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			33,294
Levelized capital charges (14.7% of total capital investment)			7,065
Total levelized annual revenue requirements			40,359

	M$	Mills/kWh
First-year annual revenue requirements	24.7	9.0
Levelized annual revenue requirements	40.4	14.7

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-75. CASE 3, 500-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	1,266
Feed preparation	2,363
Flue gas handling	11,175
SO_2 absorption	18,070
Solids separation	2,265
Total process capital	35,139
Services, utilities, and miscellaneous	2,108
Total direct investment excluding waste disposal	37,247
Waste disposal	847
Total direct investment	38,094
Indirect Investment	
Engineering design and supervision	2,607
Architect and engineering contractor	745
Construction expense	5,960
Contractor fees	1,862
Contingency	4,842
Waste disposal indirect investment	292
Total fixed investment	54,402
Other Capital Investment	
Allowance for startup and modifications	4,261
Interest during construction	8,487
Land	113
Working capital	2,108
Total capital investment	69,371
$/kW	138.7

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-76. CASE 3, 500-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Limestone	21,871 tons	8.50	186
Total raw material cost			186
Conversion costs			
Operating labor and supervision			
Process	39,610 man-hr	15.00	594
Landfill	6,036 man-hr	21.00	127
Utilities			
Process water	136,178 kgal	0.14	19
Electricity	39,914,421 kWh	0.037	1,477
Diesel fuel	17,462 gal	1.60	28
Maintenance			
Labor and material			3,005
Analysis	3,300 man-hr	21.00	69
Total conversion costs			5,319
Total direct costs			5,505
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,277
Total first-year operating and maintenance costs			7,782
Levelized capital charges (14.7% of total capital investment)			10,198
Total first-year annual revenue requirements			17,980
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			14,677
Levelized capital charges (14.7% of total capital investment)			10,198
Total levelized annual revenue requirements			24,875

	M$	Mills/kWh
First-year annual revenue requirements	18.0	6.5
Levelized annual revenue requirements	24.9	9.0

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-77. CASE 3, 500-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Particulate removal and storage	14,354
Particulate transfer	4,378
Flue gas handling	4,290
Total process capital	23,022
Services, utilities, and miscellaneous	1,381
Total direct investment excluding waste disposal	24,403
Waste disposal	2,628
Total direct investment	27,031
Indirect Investment	
Engineering design and supervision	1,708
Architect and engineering contractor	488
Construction expense	3,904
Contractor fees	1,220
Contingency	6,345
Waste disposal indirect investment	905
Total fixed investment	41,601
Other Capital Investment	
Allowance for startup and modifications	3,807
Interest during construction	6,490
Land	313
Working capital	1,335
Total capital investment	53,546
$/kW	107.1

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-78. CASE 3, 500-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	550 tons	65	36
Total raw material cost			36
Conversion costs			
Operating labor and supervision			
Process	15,330 man-hr	15.00	230
Landfill	18,710 man-hr	21.00	393
Utilities			
Process water	2,330 kgal	0.14	0
Electricity	26,830,622 kWh	0.037	993
Diesel fuel	54,129 gal	1.60	87
Maintenance			
Labor and material			1,299
Analysis	300 man-hr	21.00	6
Total conversion costs			3,008
Total direct costs			3,044
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,157
Total first-year operating and maintenance costs			4,201
Levelized capital charges (14.7% of total capital investment)			7,871
Total first-year annual revenue requirements			12,072
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			7,923
Levelized capital charges (14.7% of total capital investment)			7,871
Total levelized annual revenue requirements			15,794

	M$	Mills/kWh
First-year annual revenue requirements	12.1	4.4
Levelized annual revenue requirements	15.8	5.7

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-79. CASE 3, 500-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	15,997
SO_2 removal areas	35,139
Particulate removal areas	23,022
Total process capital	74,158
Services, utilities, and miscellaneous	4,449
Total direct investment excluding waste disposal	78,607
Waste disposal	3,505
Total direct investment	82,112
Indirect Investment	
Engineering design and supervision	5,502
Architect and engineering contractor	1,572
Construction expense	12,577
Contractor fees	3,930
Contingency	15,596
Waste disposal indirect investment	1,207
Total fixed investment	122,496
Other Capital Investment	
Allowance for startup and modifications	10,713
Interest during construction	19,110
Royalties	563
Land	441
Working capital	4,200
Catalyst	13,455
Total capital investment	170,978
$/kW	342.0

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-80. CASE 3, 500-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,165 tons	155	336
Catalyst	660 tons	23,558	15,549
Sodium hydroxide	119 tons	300	36
H_2SO_4 (100% equivalent)	610 tons	65	40
Limestone	21,871 tons	8.50	186
Total raw material cost			16,147
Conversion costs			
Operating labor and supervision			
Process	59,320 man-hr	15.00	890
Landfill	24,960 man-hr	21.00	524
Utilities			
Steam	19,190 MBtu	3.30	63
Process water	146,477 kgal	0.14	20
Electricity	77,317,503 kWh	0.037	2,861
Diesel fuel	72,211 gal	1.60	116
Maintenance			
Labor and material			4,983
Analysis	5,790 man-hr	21.00	121
Total conversion costs			9,578
Total direct costs			25,725
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,911
Total first-year operating and maintenance costs			29,636
Levelized capital charges (14.7% of total capital investment)			25,134
Total first-year annual revenue requirements			54,770
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			55,894
Levelized capital charges (14.7% of total capital investment)			25,134
Total levelized annual revenue requirements			81,028

	M$	Mills/kWh
First-year annual revenue requirements	54.8	19.9
Levelized annual revenue requirements	81.0	29.5

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-81. CASE 3, 500-MW, NO_x REMOVAL CAPITAL INVESTMENT, 90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	1,394
Reactor	9,937
Flue gas handling	5,388
Air heater	861
Total process capital	17,580
Services, utilities, and miscellaneous	1,055
Total direct investment excluding waste disposal	18,635
Waste disposal	35
Total direct investment	18,670
Indirect Investment	
Engineering design and supervision	1,304
Architect and engineering contractor	373
Construction expense	2,982
Contractor fees	932
Contingency	4,845
Waste disposal indirect investment	12
Total fixed investment	29,118
Other Capital Investment	
Allowance for startup and modifications	2,907
Interest during construction	4,542
Royalties	563
Land	22
Working capital	826
Catalyst	15,901
Total capital investment	53,879
$/kW	107.8

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-82. CASE 3, 500-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,432 tons	155	377
Catalyst	780 tons	23,558	18,375
Sodium hydroxide	119 tons	300	36
H_2SO_4 (100% equivalent)	60 tons	65	4
Total raw material cost			18,792
Conversion costs			
Operating labor and supervision			
Process	4,380 man-hr	15.00	66
Landfill	253 man-hr	21.00	5
Utilities			
Steam	20,454 MBtu	3.30	68
Process water	7,978 kgal	0.14	1
Electricity	10,703,620 kWh	0.037	396
Diesel fuel	731 gal	1.60	1
Maintenance			
Labor and material			746
Analysis	2,190 man-hr	21.00	46
Total conversion costs			1,329
Total direct costs			20,121
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			518
Total first-year operating and maintenance costs			20,639
Levelized capital charges (14.7% of total capital investment)			7,920
Total first-year annual revenue requirements			28,559
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			38,925
Levelized capital charges (14.7% of total capital investment)			7,920
Total levelized annual revenue requirements			46,845

	M$	Mills/kWh
First-year annual revenue requirements	28.6	10.4
Levelized annual revenue requirements	46.8	17.0

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-83. CASE 3, 500-MW, SO_2 REMOVAL CAPITAL INVESTMENT,

90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
Materials handling	1,267
Feed preparation	2,363
Flue gas handling	11,180
SO_2 absorption	18,072
Solids separation	2,265
Total process capital	35,147
Services, utilities, and miscellaneous	2,109
Total direct investment excluding waste disposal	37,256
Waste disposal	847
Total direct investment	38,103
Indirect Investment	
Engineering design and supervision	2,608
Architect and engineering contractor	745
Construction expense	5,961
Contractor fees	1,863
Contingency	4,843
Waste disposal indirect investment	292
Total fixed investment	54,415
Other Capital Investment	
Allowance for startup and modifications	4,262
Interest during construction	8,489
Land	113
Working capital	2,108
Total capital investment	69,387
$/kW	138.8

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-84. CASE 3, 500-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS,

90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Limestone	21,871 tons	8.50	186
Total raw material cost			186
Conversion costs			
Operating labor and supervision			
Process	39,610 man-hr	15.00	594
Landfill	6,036 man-hr	21.00	127
Utilities			
Process water	136,269 kgal	0.14	19
Electricity	39,935,729 kWh	0.037	1,478
Diesel fuel	17,462 gal	1.60	28
Maintenance			
Labor and material			3,006
Analysis	3,300 man-hr	21.00	69
Total conversion costs			5,321
Total direct costs			5,507
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,278
Total first-year operating and maintenance costs			7,785
Levelized capital charges (14.7% of total capital investment)			10,200
Total first-year annual revenue requirements			17,985
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			14,683
Levelized capital charges (14.7% of total capital investment)			10,200
Total levelized annual revenue requirements			24,883

	M$	Mills/kWh
First-year annual revenue requirements	18.0	6.5
Levelized annual revenue requirements	24.9	9.0

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-85. CASE 3, 500-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT,

90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
Particulate removal and storage	14,354
Particulate transfer	4,378
Flue gas handling	4,290
Total process capital	23,022
Services, utilities, and miscellaneous	1,381
Total direct investment excluding waste disposal	24,403
Waste disposal	2,628
Total direct investment	27,031
Indirect Investment	
Engineering design and supervision	1,708
Architect and engineering contractor	488
Construction expense	3,904
Contractor fees	1,220
Contingency	6,345
Waste disposal indirect investment	905
Total fixed investment	41,601
Other Capital Investment	
Allowance for startup and modifications	3,807
Interest during construction	6,490
Land	314
Working capital	1,335
Total capital investment	53,547
$/kW	107.1

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-86. CASE 3, 500-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	550 tons	65	36
Total raw material cost			36
Conversion costs			
Operating labor and supervision			
Process	15,330 man-hr	15.00	230
Landfill	18,710 man-hr	21.00	393
Utilities			
Process water	2,331 kgal	0.14	0
Electricity	26,854,413 kWh	0.037	994
Diesel fuel	54,129 gal	1.60	87
Maintenance			
Labor and material			1,299
Analysis	300 man-hr	21.00	6
Total conversion costs			3,009
Total direct costs			3,045
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,157
Total first-year operating and maintenance costs			4,202
Levelized capital charges (14.7% of total capital investment)			7,871
Total first-year annual revenue requirements			12,073
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			7,925
Levelized capital charges (14.7% of total capital investment)			7,871
Total levelized annual revenue requirements			15,796

	M$	Mills/kWh
First-year annual revenue requirements	12.1	4.4
Levelized annual revenue requirements	15.8	5.7

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-87. CASE 3, 500-MW, TOTAL CAPITAL INVESTMENT,

90% NO_x REMOVAL

	Capital investment, k$
Direct Investment	
NO_x removal areas	17,580
SO_2 removal areas	35,147
Particulate removal areas	23,022
Total process capital	75,749
Services, utilities, and miscellaneous	4,545
Total direct investment excluding waste disposal	80,294
Waste disposal	3,510
Total direct investment	83,804
Indirect Investment	
Engineering design and supervision	5,620
Architect and engineering contractor	1,606
Construction expense	12,847
Contractor fees	4,015
Contingency	16,033
Waste disposal indirect investment	1,209
Total fixed investment	125,134
Other Capital Investment	
Allowance for startup and modifications	10,976
Interest during construction	19,521
Royalties	563
Land	449
Working capital	4,269
Catalyst	15,901
Total capital investment	176,813
$/kW	353.6

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 90% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-88. CASE 3, 500-MW, TOTAL ANNUAL REVENUE REQUIREMENTS, 90% NO_x REMOVAL

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	2,432 tons	155	377
Catalyst	780 tons	23,558	18,375
Sodium hydroxide	119 tons	300	36
H_2SO_4 (100% equivalent)	610 tons	65	40
Limestone	21,871 tons	8.50	186
Total raw material cost			19,014
Conversion costs			
Operating labor and supervision			
Process	59,320 man-hr	15.00	890
Landfill	24,999 man-hr	21.00	525
Utilities			
Steam	20,454 MBtu	3.30	68
Process water	146,578 kgal	0.14	20
Electricity	77,493,762 kWh	0.037	2,868
Diesel fuel	72,322 gal	1.60	116
Maintenance			
Labor and material			5,051
Analysis	5,790 man-hr	21.00	121
Total conversion costs			9,659
Total direct costs			28,673
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,953
Total first-year operating and maintenance costs			32,626
Levelized capital charges (14.7% of total capital investment)			25,991
Total first-year annual revenue requirements			58,617
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			61,533
Levelized capital charges (14.7% of total capital investment)			25,991
Total levelized annual revenue requirements			87,524

	M$	Mills/kWh
First-year annual revenue requirements	58.6	21.3
Levelized annual revenue requirements	87.5	31.8

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-89. CASE 3, 1,000-MW, NO_x REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NH_3 storage and injection	2,198
Reactor	16,476
Flue gas handling	10,622
Air heater	1,694
Total process capital	30,990
Services, utilities, and miscellaneous	1,859
Total direct investment excluding waste disposal	32,849
Waste disposal	48
Total direct investment	32,897
Indirect Investment	
Engineering design and supervision	1,971
Architect and engineering contractor	328
Construction expense	4,599
Contractor fees	1,314
Contingency	8,212
Waste disposal indirect investment	16
Total fixed investment	49,337
Other Capital Investment	
Allowance for startup and modifications	4,927
Interest during construction	7,697
Royalties	1,091
Land	24
Working capital	1,402
Catalyst	26,706
Total capital investment	91,184
$/kW	91.2

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-90. CASE 3, 1,000-MW, NO_x REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	5,046 tons	155	782
Catalyst	1,310 tons	23,558	30,861
Sodium hydroxide	232 tons	300	70
H_2SO_4 (100% equivalent)	116 tons	65	8
Total raw material cost			31,721
Conversion costs			
Operating labor and supervision			
Process	6,570 man-hr	15.00	99
Landfill	292 man-hr	21.00	6
Utilities			
Steam	41,257 MBtu	3.30	136
Process water	15,502 kgal	0.14	2
Electricity	20,951,980 kWh	0.037	755
Diesel fuel	1,064 gal	1.60	2
Maintenance			
Labor and material			987
Analysis	2,628 man-hr	21.00	55
Total conversion costs			2,062
Total direct costs			33,783
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			688
Total first-year operating and maintenance costs			34,471
Levelized capital charges (14.7% of total capital investment)			13,404
Total first-year annual revenue requirements			47,875
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			65,012
Levelized capital charges (14.7% of total capital investment)			13,404
Total levelized annual revenue requirements			78,416

	M$	Mills/kWh
First-year annual revenue requirements	47.9	8.7
Levelized annual revenue requirements	78.4	14.3

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-91. CASE 3, 1,000-MW, SO_2 REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Materials handling	1,345
Feed preparation	2,828
Flue gas handling	21,686
SO_2 absorption	35,584
Solids separation	2,579
Total process capital	64,022
Services, utilities, and miscellaneous	3,841
Total direct investment excluding waste disposal	67,863
Waste disposal	1,306
Total direct investment	69,169
Indirect Investment	
Engineering design and supervision	4,072
Architect and engineering contractor	679
Construction expense	9,501
Contractor fees	2,715
Contingency	8,483
Waste disposal indirect investment	424
Total fixed investment	95,043
Other Capital Investment	
Allowance for startup and modifications	7,465
Interest during construction	14,827
Land	167
Working capital	3,627
Total capital investment	121,129
$/kW	121.1

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-92. CASE 3, 1,000-MW, SO_2 REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Limestone	42,433 tons	8.50	361
Total raw material cost			361
Conversion costs			
Operating labor and supervision			
Process	49,330 man-hr	15.00	740
Landfill	7,910 man-hr	21.00	166
Utilities			
Process water	264,862 kgal	0.14	37
Electricity	75,892,554 kWh	0.037	2,808
Diesel fuel	28,828 gal	1.60	46
Maintenance			
Labor and material			4,790
Analysis	4,940 man-hr	21.00	104
Total conversion costs			8,691
Total direct costs			9,052
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,480
Total first-year operating and maintenance costs			12,532
Levelized capital charges (14.7% of total capital investment)			17,806
Total first-year annual revenue requirements			30,338
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			23,635
Levelized capital charges (14.7% of total capital investment)			17,806
Total levelized annual revenue requirements			41,441

	M$	Mills/kWh
First-year annual revenue requirements	30.3	5.5
Levelized annual revenue requirements	41.4	7.5

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-93. CASE 3, 1,000-MW, PARTICULATE REMOVAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
Particulate removal and storage	27,916
Particulate transfer	6,403
Flue gas handling	8,456
Total process capital	42,775
Services, utilities, and miscellaneous	2,567
Total direct investment excluding waste disposal	45,342
Waste disposal	4,143
Total direct investment	49,485
Indirect Investment	
Engineering design and supervision	2,721
Architect and engineering contractor	453
Construction expense	6,348
Contractor fees	1,814
Contingency	11,336
Waste disposal indirect investment	1,344
Total fixed investment	73,501
Other Capital Investment	
Allowance for startup and modifications	6,801
Interest during construction	11,466
Land	490
Working capital	2,310
Total capital investment	94,568
$/kW	94.6

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-94. CASE 3, 1,000-MW, PARTICULATE REMOVAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
H_2SO_4 (100% equivalent)	1,068 tons	65	69
Total raw material cost			69
Conversion costs			
Operating labor and supervision			
Process	21,900 man-hr	15.00	329
Landfill	25,078 man-hr	21.00	527
Utilities			
Process water	4,526 kgal	0.14	1
Electricity	52,163,646 kWh	0.037	1,930
Diesel fuel	91,398 gal	1.60	146
Maintenance			
Labor and material			1,938
Analysis	400 man-hr	21.00	8
Total conversion costs			4,879
Total direct costs			4,948
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,681
Total first-year operating and maintenance costs			6,629
Levelized capital charges (14.7% of total capital investment)			13,901
Total first-year annual revenue requirements			20,530
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			12,502
Levelized capital charges (14.7% of total capital investment)			13,901
Total levelized annual revenue requirements			26,403

	M$	Mills/kWh
First-year annual revenue requirements	20.5	3.7
Levelized annual revenue requirements	26.4	4.8

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.

TABLE A-95. CASE 3, 1,000-MW, TOTAL CAPITAL INVESTMENT

	Capital investment, k$
Direct Investment	
NO_x removal areas	30,990
SO_2 removal areas	64,022
Particulate removal areas	42,775
Total process capital	137,787
Services, utilities, and miscellaneous	8,267
Total direct investment excluding waste disposal	146,054
Waste disposal	5,497
Total direct investment	151,551
Indirect Investment	
Engineering design and supervision	8,764
Architect and engineering contractor	1,460
Construction expense	20,448
Contractor fees	5,843
Contingency	28,031
Waste disposal indirect investment	1,784
Total fixed investment	217,881
Other Capital Investment	
Allowance for startup and modifications	19,193
Interest during construction	33,990
Royalties	1,091
Land	681
Working capital	7,339
Catalyst	26,706
Total capital investment	306,881
$/kW	306.9

Basis: 0.7% sulfur subbituminous coal, new pulverized-coal-fired power unit with a 30-yr life at 5,500-hr/yr full-load operation, 80% SCR NO_x reduction from 1979 NSPS level, natural-oxidation limestone FGD and hot-side ESP to meet 1979 NSPS, mid-1982 costs.

TABLE A-96. CASE 3, 1,000-MW, TOTAL ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Cost - First Year			
Ammonia	5,046 tons	155	782
Catalyst	1,310 tons	23,558	30,861
Sodium hydroxide	232 tons	300	70
H_2SO_4 (100% equivalent)	1,185 tons	65	77
Limestone	42,433 tons	8.50	361
Total raw material cost			32,151
Conversion costs			
Operating labor and supervision			
Process	77,800 man-hr	15.00	1,168
Landfill	33,280 man-hr	21.00	699
Utilities			
Steam	41,257 MBtu	3.30	136
Process water	284,890 kgal	0.14	40
Electricity	149,008,180 kWh	0.037	5,513
Diesel fuel	121,290 gal	1.60	194
Maintenance			
Labor and material			7,715
Analysis	7,968 man-hr	21.00	167
Total conversion costs			15,632
Total direct costs			47,783
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			5,849
Total first-year operating and maintenance costs			53,632
Levelized capital charges (14.7% of total capital investment)			45,111
Total first-year annual revenue requirements			98,743
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			101,149
Levelized capital charges (14.7% of total capital investment)			45,111
Total levelized annual revenue requirements			146,260

	M$	Mills/kWh
First-year annual revenue requirements	98.7	18.0
Levelized annual revenue requirements	146.3	26.6

Basis: One year of operation at the conditions described on the capital investment table, mid-1984 costs.